THE CANADIAN WATER
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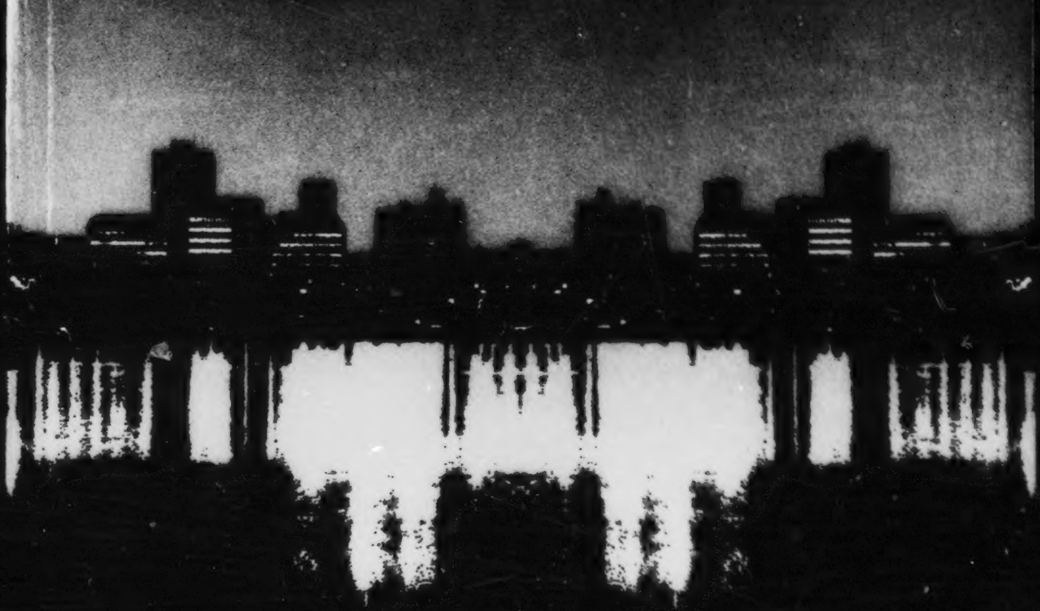

PRESENT

WHAT'S NEW IN WASTEWATER TECHNOLOGY?

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WHAT'S NEW IN WASTEWATER TECHNOLOGY?

Background papers for a Technology Transfer Seminar Series
sponsored jointly by

Environment Canada

and the

Canadian Water and Wastewater Association

1988

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CANADIAN WATER AND WASTEWATER ASSOCIATION
ASSOCIATION CANADIENNE DES EAUX POTABLES ET USÉES

As the public and governments alike become increasingly concerned about environmental protection generally and about improving water resource management in particular, Canada's municipalities are facing growing demands to introduce, renew and upgrade wastewater treatment systems and facilities.

To date, however, the increased concern about environmental protection has not been matched by increased funding. On the contrary, fiscal restraint can be seen as a competing and equally compelling concern on the part of both the public and governments. Thus it continues to be essential to meet both objectives by finding better ways of doing more things at a lower cost.

Both Environment Canada and the Canadian Water and Wastewater Association believe that the technologies developed at the Wastewater Technology Centre have the potential to provide innovative, cost-effective solutions to many wastewater treatment problems. And because one of the chief objectives of the Canadian Water and Wastewater Association is to facilitate the exchange of information among those working in the water and wastewater industries, we are pleased to be co-sponsors of the series entitled "What's New in Wastewater Technology?"

The papers contained in this publication form the basis for the presentations made at the seminars held in November, 1988, and the spring of 1989. Together with the information presented at the sessions, they provide a broad appreciation of the municipal wastewater technology being developed at the Wastewater Technology Centre in Burlington, Ontario as of the fall of 1988.

The Canadian Water and Wastewater Association also wishes to thank its sister and affiliate water and wastewater associations from across Canada for having helped to organize the sessions. Those serving as participating hosts for the series were: the Atlantic Canada Section of the American Water Works Association, the British Columbia Water and Waste Association, L'Association québecoise des techniques de l'eau, the Maritime Provinces Water and Wastewater Association, the Pollution Control Association of Ontario and the Western Canada Pollution Control Association.

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INTRODUCTION

by

B.E. Jank

Director, Wastewater Technology Centre
Environment Canada

The escalating cost of upgrading, maintaining and operating the existing municipal wastewater treatment infrastructure, expanding it to support residential and industrial growth, and providing a new infrastructure where it does not exist is creating a serious problem for Canadian municipalities. The requirement for the provision of a treatment facility that will reduce or eliminate toxic chemicals, in addition to the conventional pollutants, further complicates this issue. Toxic chemicals control also dictates that the treatment, handling and disposal of treatment process sludges or residues receive considerably more attention than in the past.

The solutions to these problems have become understandably more complex. Recent advances in reactor design, computer control technology and analytical capability have provided the tools for research scientists and engineers to develop cost-effective solutions in many of these cases.

Environment Canada's Wastewater Technology Centre has been a major contributor to the team of national and international researchers that has developed and continues to develop appropriate solutions for municipal and industrial environmental problems. The emphasis has been on the optimization of the design and operation of existing wastewater treatment systems. Although this approach has resulted in the development of some new technology, most of the advancements have been achieved through upgrading and modifying existing unit operations.

For industries discharging to a municipal sewer system, product re-use, generation of alternative products and modifications to existing processes represent the essential initial components of an effective municipal waste management program. If source control is essential, metal recovery systems or energy efficient high-rate anaerobic systems should be considered. If hazardous or toxic wastes are present which cannot be eliminated, recovered or treated at an on-site facility, then these wastes must be transferred to an appropriately licensed centralized treatment, handling and disposal facility. The municipality should ensure that these source control programs are in place and effectively controlled. The impact of industrial waste source control on the hydraulic and organic loading capacity and operating complexity of the municipal wastewater treatment facility will be very significant, eventually resulting in substantial savings in capital and operating costs.

An excellent example of cost-effective industrial waste source control is the application of high rate anaerobic technology for the pretreatment of high organic strength, high temperature wastes. The methane generated from the anaerobic process located at the industrial plant will be used by the industry to offset the purchase of natural gas or fuel oil. In most cases, the percentage reduction of organics will be sufficient to provide an effluent that will meet municipal industrial pretreatment requirements, and will eliminate or significantly reduce industrial waste discharge surcharges. Significant energy savings will accrue at the central treatment facility as a result of reduced aeration requirements and corresponding reductions in sludge production. Dairy, brewery, distillery, slaughter house and fruit and vegetable processing wastewaters represent excellent examples where high-rate anaerobic technology can be used at source to reduce the total cost of wastewater treatment for both the industry and the municipality.

Major advances in the optimization of the design and operation of municipal wastewater treatment plants have resulted from automated process control. These include techniques for optimizing the efficiency of the aeration equipment and the design and operation of the secondary clarifier, which both have the potential of providing substantial capital and operation and maintenance savings. On-line instrumentation coupled to computers programmed with appropriate automated control strategies will improve the operation of the activated sludge process and many other unit operations within the treatment system. The computer control procedures have been used to develop a process audit to assess the operational performance of full scale wastewater treatment plants. The process audit can be used to identify design deficiencies, the anticipated life of the facility (following the correction of design deficiencies), the specifications for the computer control system and the potential energy savings following incorporation of computer control.

Results of process audits conducted to date have been used to determine that it may be possible to reduce projected infrastructure upgrading costs of $1.3 billion by $1 billion if design deficiencies are corrected and a computer control system is installed, rather than expanding the capacity of the plant. The results emphasize the importance of a process audit prior to proceeding with the upgrading or expansion of an existing wastewater treatment plant.

Properly designed and operated municipal treatment facilities have been shown to produce an effluent having an acceptable level of trace organics and heavy metals. Computer controlled facilities will provide consistent effluent quality, an important factor for those plants designed to provide toxic chemicals control. Appropriate sludge treatment, handling and disposal facilities must be included in the treatment facility. This is an area that has not previously received the necessary degree of attention from the design and operations perspective.

For economic and technical reasons, whenever possible, land application of sludge should be the selected disposal option. Land application guidelines have been prepared for Canada and other developed countries and it is evident that the environmental risk associated with land application is minimal. Anaerobic sludge digestion is frequently an integral component of the land application option. Improvement of digestion mixing by optimizing the number of mixing devices and the mixing energy could extend the life of the existing digester facilities at a fraction of the capital cost of providing new digesters.

Approximately 50 percent of the capital and operating costs of the municipal wastewater treatment facility are required for the sludge management system. The introduction of innovative technologies such as the WRC (Water Research Centre) sludge thickener, the CRIQ (Centre de recherche industrielle du Québec) ring press, autothermal thermophilic aerobic digestion and Environment Canada's oil-from-sludge technology will result in cost reductions for specific applications. The development of a control system such as the sludge conditioning controller developed by the Wastewater Technology Centre and licensed to Zenon Environmental Inc. in Burlington will substantially reduce operating costs. It will also force plant owners to take a close look at their entire sludge management train, as virtually every system requires upgrading and improved operation.

It is obvious that on-line computer control represents an important component of the upgrading program. Extensive experience has shown that the computer hardware and software is available, affordable and reliable. The weak link in the system is still the primary sensor. To improve the quality of existing sensors and their reliability, the Wastewater Technology Centre has participated in the formation of the Water and Wastewater Instrumentation Testing Association of North America (ITA). ITA prepares detailed test protocols used for bench to field testing of commercially available instrumentation. Members receive sensor evaluation reports and copies of the protocol which can be used for tender specifications. They also have access to an instrumentation clearing house established to provide long-term operating experience for as yet untested primary sensors. The cost of instrumentation specification and partial testing for procurement dictates that virtually every municipality having a computer controlled treatment facility should be a member of ITA.

Environment Canada's Wastewater Technology Centre has been extensively involved in the development, design and operation of the process technology and automated control systems for industrial pretreatment and municipal wastewater treatment. This CWWA/Environment Canada technology transfer seminar has been organized to assist in ensuring that information from these programs is disseminated to municipalities and their consultants across Canada.

FIELD VERIFICATION OF ON-LINE INSTRUMENTATION AT A MUNICIPAL WASTEWATER TREATMENT PLANT

by

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Abstract

As part of a full-scale project to demonstrate the advantages of using automated process control in wastewater treatment plants, on-line sensors used in various control loops were calibrated and/or verified during control system commissioning and startup. The on-line measurements included liquid flow rates, air flow rates, suspended solids concentrations, and dissolved oxygen concentrations. The methods used to confirm instrument performance are presented along with results that show that several instruments were found to operate outside the manufacturers' written specifications. In addition, on-line methods which can help determine when recalibrations are required, and thereby help to reduce maintenance costs and ensure reliability, are also presented.

Keywords

Activated sludge, aeration, calibration, dissolved oxygen, flow measurement, instrumentation, suspended solids, process control.

Introduction

Environment Canada's Wastewater Technology Centre (WTC) initiated a project in 1982 to demonstrate that energy savings and improved operation of municipal wastewater treatment plants (WWTP) result from the use of direct digital control strategies. The Tillsonburg, Ontario, activated sludge plant was ultimately selected for the study in 1983. As part of the selection process, an operational audit was conducted utilizing on-line instrumentation coupled to a microcomputer to collect dynamic data on the plant's loading and operating characteristics (Speirs and Stephenson, 1985). An estimated 30 to 40 percent saving in aeration energy was predicted with automated aeration control. Refit modifications were started for the segregation of the plant into parallel process trains for comparison of manual and automated control. Data from the operational audit were used as the basis for these design modifications. The refit of the plant was completed in October 1984.

The on-line instrumentation and computer control system for the demonstration was installed by WTC staff. An essential part of the control system start-up consisted of the field calibration and/or verification of all the critical instruments and final control elements in each loop. This paper outlines the field calibration techniques and verification methods utilized. The control loops investigated included recycle sludge control, waste sludge flow rate, and aeration. The instruments included liquid flow meters, air flow meters, suspended solids meters, dissolved oxygen meters, and several other commonly used transmitters.

Plant Description

The Tillsonburg WWTP provides conventional activated sludge treatment of municipal and commercial wastewaters for the Town of Tillsonburg, Ontario. The plant was originally constructed in 1960 and expanded in 1972. The expansion doubled the plant capacity to 9000 m^3/day. Essentially all major unit operations and associated machinery were duplicated. This resulted in the plant having two equally sized, parallel trains, herein labelled north and south. The plant was refit to ensure that the process trains were segregated. Each train consisted of a primary settling tank, two aeration tanks-in-series, a final settling tank, and a chlorine contact tank. The aeration system for each train consisted of one 45 kW and one 30 kW positive displacement (PD) blower discharging into a common header which ultimately split into parallel 150 mm (6 in.) and 100 mm (4 in.) distribution piping for air delivery to the first and second aeration passes respectively. For the south train, the air flow rate and distribution were controlled manually while for the north train they were controlled automatically. A schematic of the plant is shown in Figure 1.

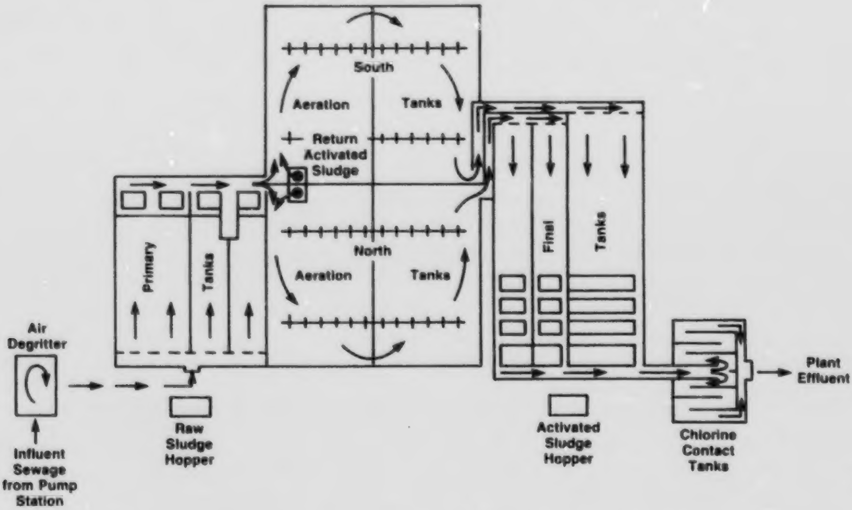

Fig. 1. Schematic of Tillsonburg WWTP tankage (not to scale)

The process control computer chosen for the project was a Hewlett-Packard (HP) 1000 series A-600 minicomputer with 1 megabyte of memory. This computer was selected because of previous WTC experience with similar models and the extensive library of existing software which had been developed for other projects at the WTC. The HP 1000 computer is classified as a true real-time, multi-tasking, multi-user system which can run many "simultaneous" tasks and handle multiple users at the same time. Due to the research nature of the tasks undertaken at the plant, data acquisition and control software was developed in-house by WTC programmers and engineers. It was considered that most commercially available software packages for data acquisition and control were too limited for this specific application.

On-line instrumentation was installed to monitor the performance of both process trains. The sensors also provided data upon which automatic control actions were initiated by the computer. A summary of the installed instrumentation is provided in Table 1. In addition, current detect relays were used to provide positive feedback of pump/motor operating status. Final control elements included air distribution control valves, a.c. variable frequency drives for blower and recycle pump control, and relays for on/off control of pumps, motors, and valves. As part of commissioning the monitoring and process control system, all on-line sensors were calibrated and their accuracy verified. The following paragraphs describe the techniques used during the calibration/verification stage and the results obtained.

Table 1. Tillsonburg On-Line Instrumentation

VARIABLE	LOCATION	INSTRUMENT
DO	Aeration basins	Polarographic probe
Flow Rate	Waste sludge	Magnetic flow meter
Level	Effluent flow	Capacitance probe
	Recycle flow	Ultrasonic detector
	Pump station wet well	Bubbler (pressure)
Turbidity	Effluent	Direct Reading Turb.
Suspended Solids	MLSS, RAS, Primary effluent	Optical probe
Air Flow Rate	Each aeration basin	Vortex or Orifice meter
Power (watts,vars)	Blowers	Transducer
Temperature	Air lines, MLSS, Ambient	RTD element
Pressure	Air lines, Atmospheric	Transducer
Humidity	Ambient	R.H. meter
a.c. Current	Motors (pumps, blowers)	Contact relay

Discussion of Instrument Calibration/Verification Techniques

Return Sludge Measurement

The primary variable of interest in the return sludge control system was the recycle flow rate itself. The only available point of measurement was at an existing weir box through which the return activated sludge (RAS) was recycled to the aeration tanks. This was accomplished by measuring the liquid height over the 90 degree V-notch weir with an ultrasonic level detector. The weir box itself was very small with limited approach length and a large amount of surface turbulence. This was not equivalent to the "ideal" conditions assumed in most weir equations and potentially could have limited the accuracy of the level measurement. The surface turbulence was reduced to an acceptable level by installing a baffle over the inlet pipe in the weir box, however, the approach velocity of the sludge was quite significant. Therefore, it was necessary to obtain a number of flow rates and corresponding liquid heights to confirm the applicability of the theoretical weir equation or to derive a more appropriate relationship.

There are several methods commonly used for measuring flow rates. With the dilution method, a known quantity of dye (or any conservative tracer) is continuously injected into the stream, and the resulting dye concentration is measured after thorough mixing (Turner Designs). By knowing the rate of addition of dye and the resulting concentration, the flow rate may be calculated. The dilution method may be used in many situations but requires the use of a fluorescence spectrophotometer and a very accurate metering pump. Another common method used for flow verification is the draw and fill method. In this procedure, the level of liquid in a tank is measured as it is being emptied or filled. The flow rate is the product of the rate of change of the water level and the surface area of the tank. This methodology is applicable in many wastewater treatment plants where a reservoir such as a hopper or extra tank can be isolated. Accurate timing and simultaneous reading of tank level and flow meter output is required.

At the Tillsonburg plant the draw and fill method was utilized because an acceptable reservoir was available. The RAS hopper was isolated from the clarifier by means of slip pipe valves (see Figure 2). The water level in the hopper was measured with an ultrasonic level detector temporarily mounted above the hopper. A computer program automatically collected both the RAS hopper level and the discharge weir liquid height on a five second basis. After completion of a run, the RAS hopper level data was differentiated numerically (Hershey et al., 1967) and the flow rate calculated. Figure 3 shows a comparison of the calculated and the measured flow rate based upon the ideal weir equation.

The results of a series of draw down tests at several different flow rates showed good agreement between the measured and the estimated flow rates. Since the tests involved several measurements, each with an associated error , it was concluded that the use of the existing weir box and the theoretical 90 degree v-notch weir equation resulted in accurate measurements even under the non-ideal conditions.

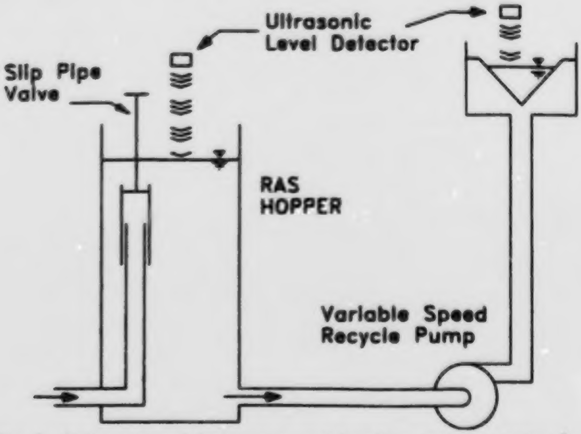

Fig. 2. Schematic of hardware used for flow meter verification

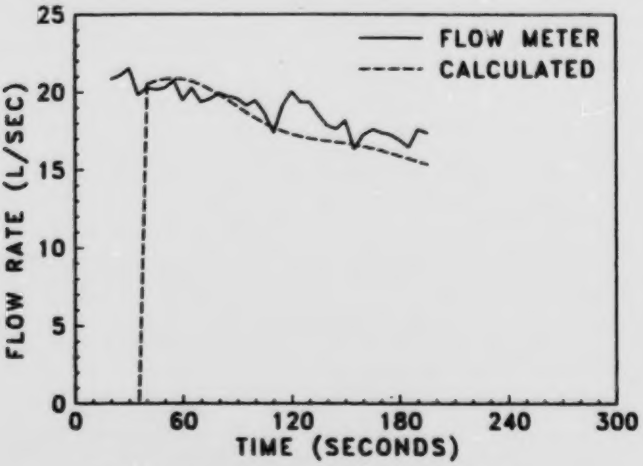

Fig. 3. Verification results for RAS flow rate

Sludge Wasting Measurement

A solids retention time (SRT or sludge age) algorithm was implemented to control the amount of solids in the activated sludge system to ensure adequate treatment. An additional benefit of the precise control of SRT was that process performance criteria, such as nitrification, could usually be maintained in the automatically controlled train at a lower sludge age than in the manually controlled side of the treatment system. This lower sludge age resulted in less oxygen demand and a power savings in aeration as well as increased digester gas production (Stenstrom and Andrews, 1980).

By definition, the sludge age or SRT of the system is the mass of aeration tank solids divided by the mass of solids wasted, both intentionally and unintentionally, and is usually expressed in terms of days. This is shown in Equation 1.

$$SRT = \frac{V \cdot MLSS}{F_E \cdot X_E + F_W \cdot RASS} \qquad (1)$$

Where

SRT = solids retention time,
V = volume of aeration basins,
MLSS = mixed liquor suspended solids concentration,
F_E = effluent flow rate,
X_E = effluent suspended solids concentration,
F_W = intentional waste sludge flow rate,
RASS = recycle/waste activated sludge solids concentration; assumes wasting from the recycle sludge handling train.

As shown in Equation 1, the variables of interest are the mixed liquor suspended solids (for calculating system solids), the effluent and waste sludge flow rates, and the effluent and waste sludge suspended solids concentration. At the Tillsonburg plant, the effluent flow rates for the south and north trains are measured with a rectangular weir and Parshall flume respectively. Calibration of these devices was straightforward and will not be discussed further. Since the effluent solids concentration was usually rather low (<10 mg/l), this term was small. The effluent solids measurement therefore had limited impact on the sludge wasting controller and although included to ensure accuracy, will not be discussed in detail. Therefore, the main measurements of interest were the waste sludge flow rate, the waste sludge suspended solids concentration, and the MLSS concentration.

The waste sludge flow rates in both trains were measured by a.c. magnetic flow meters. These meters were of different manufacture, over ten years old and utilized simply because of their availability. Due to their age, both were sent to their respective manufacturers for maintenance and calibration before installation at the plant. Once in place, both were field tested using the same draw down technique utilized to confirm the recycle flow measurements.

Figure 4a shows quite acceptable results for the waste flow meter on the automatically controlled train. However, the measurement on the manually controlled train showed a discrepancy of approximately fifteen percent between the meter output and the calculated flow (Figure 4b). Further tests at the WTC determined that this error was caused by the varying operating voltage at the plant. When the input voltage was over 120 V a.c., the meter output at a constant flow rate was less than that at 120 V a.c. Since the line voltage at the Tillsonburg plant frequently varied in the 115-135 V a.c. range, the meter was replaced with a newer d.c. magnetic flow meter, rather than provide input voltage regulation. If this error had not been found, the operator would have, for example, at a given set of operating conditions, actually maintained a SRT of 5.3 days when a SRT of 6 days was desired.

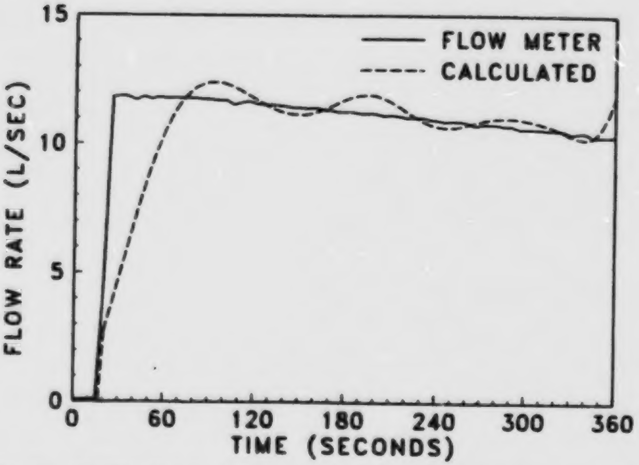

Fig. 4a. Verification results for north train WAS magnetic flow meter

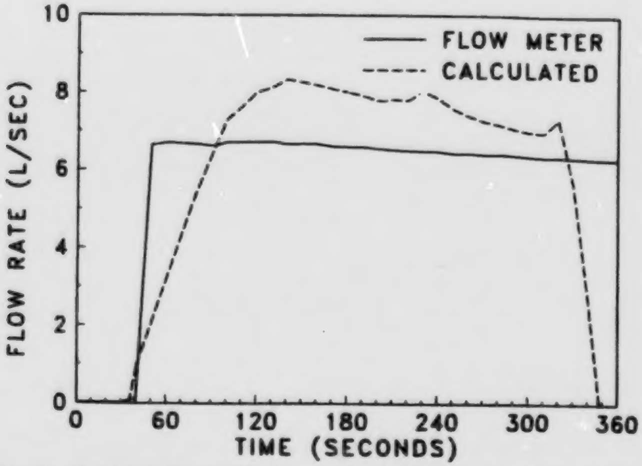

Fig. 4b. Verification results for south train WAS magnetic flow meter

The other variables of interest in the waste sludge control loop were the MLSS and RAS concentrations. Both were measured with an optical, turbidity reading instrument. As with all other on-line instruments used in the project, these instruments were initially ranged and calibrated by comparing the instrument output as read at the computer in millivolts (mV) with reference values for a number of samples. Data pairs consisting of a reference reading and the corresponding digital output from the test sensor were collected over the range of the instrument. The data pairs were plotted to determine the relationship between the meter output and reference values. When the digital values varied linearly with the reference readings, a calibration equation based on simple linear regression was used for conversion of instrument signals to engineering values in the data acquisition software. For a relationship between the digital values and reference readings which was non-linear, a curve or combination of lines and/or curves was fitted to the data. The details of this technique have been reported by Speirs et al. (1982).

Typically, for the sludge meters, the relationship was linear. Once an instrument was ranged, the meter transmitter adjustments (span and zero) were seldom touched. Changes in the true calibration due to variations in sludge characteristics were handled in the data acquisition software by re-establishing the model parameters of the relationship developed as outlined above. An added benefit of this technique was that since the hardware was not being manually adjusted, changes in the calibration could be tracked and used as an indication of long term instrument integrity.

Since complete recalibrations were quite time consuming, they were initiated only when periodic grab sample gravimetric analysis showed consistent disagreement with the meter output. To provide a more continuous indication of conformance, a technique was developed to compare the measured MLSS concentration with a predicted MLSS concentration to determine when such recalibrations were necessary. The predicted value was estimated from a mass balance around the aeration basins using on-line measurements of influent flow rate, recycle flow rate, and RAS concentration.

Under normal conditions, good agreement was observed between the measured and predicted values. However, a divergence of the values indicated a problem with one or more of the variables used in the mass balance equations. Usually, it was found that one of the suspended solids meters was out of calibration. Figure 5 is an example of one such divergence caused by the failure of the power supply of the RAS solids meter.

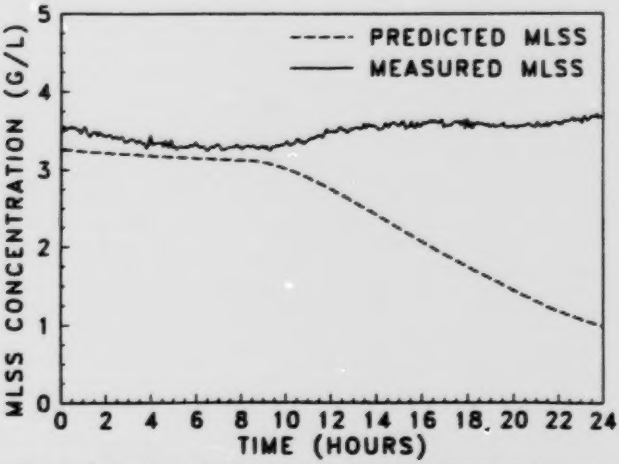

Fig. 5. Indication of instrument failure using predicted versus measured data for MLSS concentration

It is important to note that while the use of such on-line techniques is proposed as an aid to the operator, they provide information which must be used in conjunction with other conformance checks. It is quite conceivable that in the technique outlined above, inaccuracies of more than one instrument could offset each other and be masked in the mass balance equation used for the prediction. However, when used with periodic checking of on-line instruments with off-line reference determinations, these techniques are powerful tools. At the Tillsonburg plant, the use of two solids meters, each measuring in a different range, coupled with a mass balance, was adequate to determine most failures. Other procedures, using hydraulic and settler models for comparing measured and predicted values of liquid flow rates, suspended solids, and clarifier performance parameters to indicate instrument failure or changes in plant operation, are presented by Olsson *et al*. (1984).

Dissolved Oxygen Measurement

The primary objective of the Tillsonburg project was to demonstrate automated control of the dissolved oxygen concentration (DO) in the two aeration basins of the north train while minimizing aeration energy consumption. The control system consisted of a number of cascaded feedback control loops as well as several optimization routines. The control system has been fully described elsewhere (Speirs *et al*., 1985). The primary measurements in the control system included DO, air flow rate, and air header temperature and pressure. The temperature and pressure measurements will not be discussed.

Polarographic probes were utilized for DO measurements. These probes incorporated a disposable and easily replaceable measuring cell. Each probe was inserted through a round float such that the membrane of the electrochemical cell was located flush with the surface. This helped prevent hair and rags from hanging up on the probe and increased the velocity of the water at the membrane surface to minimize biofilm fouling.

In evaluating the performance of the probe, linearity was first investigated by comparing DO meter output to the reading from a calibrated Yellow Springs Instrument Company (YSI) DO meter which was used as a reference. Using the technique outlined above, it was determined that the probe was linear over the entire range and that a two point determination was adequate for most calibrations.

As with all instruments at the plant, the hardware adjustments (span and zero) of the DO meters were not changed once an acceptable range was established. Changes in calibration were handled by modifying software conversion parameters only. Therefore, the long-term stability of the instruments could be checked by inspecting the changes in calibration equation parameters with time.

Figure 6 illustrates the stability of the DO sensor as determined by seven calibration checks over a 20-month period. Graphically, it appeared that the periodic calibration of the instruments showed consistency in meter performance. Numerically, the initial calibration equation, determined on November 16, 1983, was used to determine that for reference YSI readings of 2.0 and 6.0 mg/L, the corresponding field instrument outputs measured in mV at the computer were 3079 and 5092 respectively. For each of the subsequently determined calibration equations, Figure 7 displays the YSI readings which correspond to outputs of 3079 and 5092 mV from the field sensor. Perfect agreement, i.e., YSI reference readings of 2.0 mg/L and 6.0 mg/L corresponding to 3079 and 5092 mV for each calibration over the entire twenty month period, would mean that the output of the field instrument had not changed. As seen from Figure 7, field meter performance was very consistent over the time period, as the YSI values corresponding to meter outputs of 3079 and 5092 mV did not vary by more that +/- 0.5 mg/L from the initial calibration values. These results are particularily impressive considering that the individual calibrations were conducted under variable seasonal conditions by different personnel using wastewater, clean water, air saturated water and saturated air calibration techniques. In addition, the number of data points used to generate the calibration equation for the seven cases varied from two to twenty three. Overall, the performance and reliability of the DO monitoring system was excellent.

Air Flow Measurement

Air flow measurement was originally accomplished with vortex shedding air flow meters. These units were selected for the full scale project based upon their low head loss

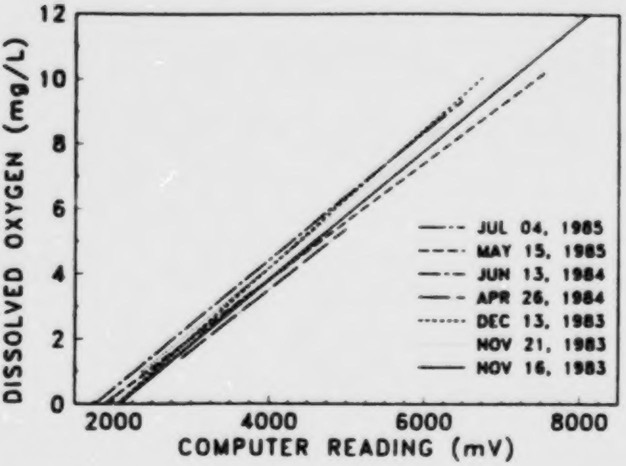

Fig. 6. Dissolved oxygen sensor calibration data over a 20-month period

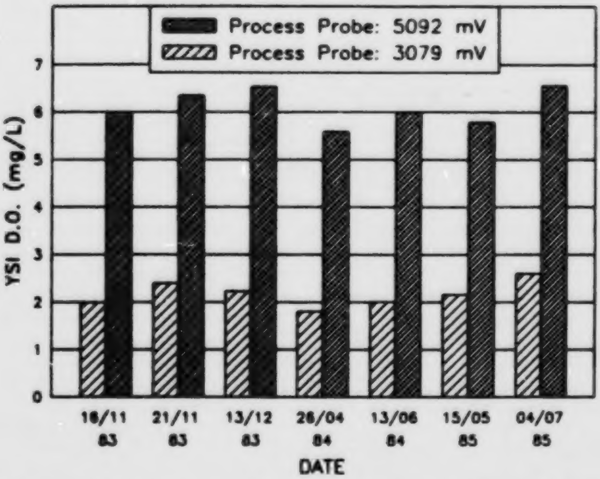

Fig. 7. Dissolved oxygen sensor performance over a 20-month period

requirements and because they had been successfully used for an identical but smaller PD blower-based air delivery system on a package scale activated sludge plant at the WTC. The manufacturer ran wet tests on each meter. Field verification was accomplished by routing all the air from one blower through each meter and comparing the meter output with the air flow calculated from the blower manufacturer's performance curves. By varying the blower speed, the entire range of the meter was tested.

Figure 8a shows the results obtained for one of the 100 mm vortex meters using the 30 kW blower. All flow rates are in the units of actual litres per second at the pressure and temperature conditions present at the meter. The dashed line on the graph represents perfect agreement between the measured and calculated flows over the expected operating range of the meter. The data shows a very good correlation between the values, especially considering that the calculated flow rate involves four measurements (ambient temperature, blower speed, air temperature, and air pressure), and the interpolation of the blower performance curves.

Results from the verification tests on the north 150-mm vortex meter are shown in Figure 8b. Again, the dashed line represents perfect agreement between calculated and measured values and extends over the expected range of operation of the meter. Large discrepancies were noted between the calculated and measured values at the lower air flow rates delivered by the 30-kW blower. In fact, when receiving approximately 125 litres per second, the vortex meter showed no response at all.

The problem with the 150-mm vortex meters was examined over a long time period. The manufacturer provided a complete extra sensor and alternate power supply for use in exploring possible grounding problems. Both meters and transmitters were interchanged between the south side, where they sometimes worked, and the north side, where they never worked. To assess the potential grounding problem, a 24-volt battery was used for the power supply. None of the corrective actions resulted in better performance in the north train. Ultimately, it was concluded that in this particular application, the interaction between the pulsing air supply delivered by the 30-kW positive displacement blower and the vortex counting transducer resulted in unreliable performance of the 150-mm vortex meter.

This problem was not evident in either the north or the south trains when the larger 45-kW blower, which operated at a slower speed and delivered larger pulses of air at a different frequency, was used. In addition, the airflow meters in the north train were located considerably closer to the blowers than the meters in the south train. It was postulated that this resulted in differing degrees of "attenuation" of the pulsing air supply by the distribution piping.

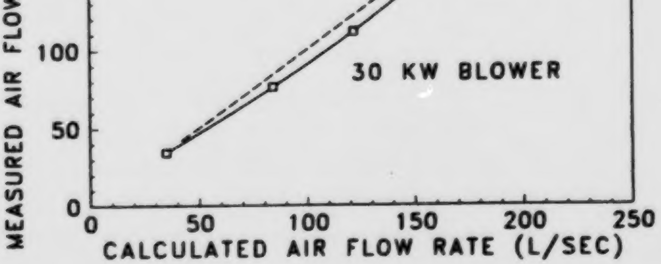

Fig. 8a. Verification results for south train 100 mm vortex air flow meter

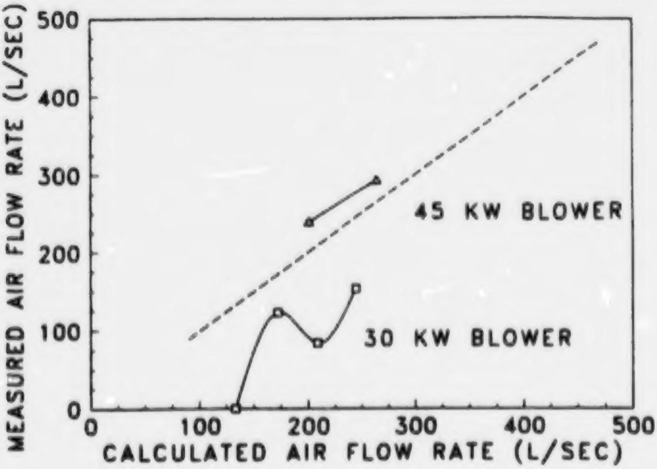

Fig. 8b. Verification results for north train 150-mm vortex air flow meter

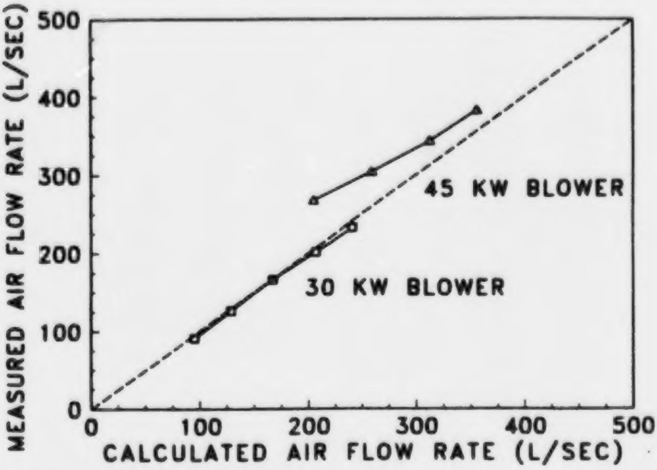

Fig. 8c. Verification of 150-mm orifice plate for air flow measurement

As an alternative to the vortex air flow meters, orifice plates were tested. The results of the orifice plate testing in the north 150-mm line were quite impressive (Figure 8c) for the low air flow rates discharged by the 30-kW blower. For the 45-kW blower, the measured air flow rate exceeded that estimated from the blower curves which was consistent with results using the vortex meter. This may have resulted from operating the 45-kW blower below the range recommended by the manufacturer and, as a consequence, having to extrapolate the blower curves below their limits as provided by the manufacturer. Agreement was better where air flow rates were within the normal operating range of the blower. It was therefore concluded that the air flow measurement provided by the orifice plate was accurate.

Summary

The experiences obtained during the start-up phase of the demonstration project at Tillsonburg confirm the need to verify the accuracy of on-line instruments installed in wastewater treatment plants. Most of the instruments, including those recently purchased, required field calibration to produce accurate results. In addition, several of the instruments were found to operate outside the written specifications provided by the manufacturer.

For the case of the magnetic flow meters used in the waste sludge line, variations in the a.c. supply voltage at the Tillsonburg plant had a large influence on the performance of one of the meters. In the case of the vortex meters used for air flow measurement, although the technology had been observed to operate successfully for an identical but smaller application at the WTC, at Tillsonburg, extensive field checking established that the particular unit supplied was not capable of producing a stable and accurate measurement under certain conditions.

As seen in these examples, it is completely unacceptable to place blind faith in the measurements obtained from on-line sensors which have not been verified. Frequently, the specifications provided by instrument manufacturers relate to "electronic" calibrations which verify the integrity of the transmitter electronics only. Even when an instrument has been fully "bench" or "wet" tested, it is still important to check instrument accuracy when installed in the process and interfaced to a computer or controller. Only when all elements of the monitoring and control loop are connected with the sensor, can the user be sure that the system is being checked under actual field conditions.

In addition to providing initial calibration and verification, all instruments require periodic recalibration and maintenance. Maintenance procedures are usually based on a combination of the recommendations provided by the instrument manufacturer and the user's experience with the unit. Most field calibration techniques are site specific and must be carefully considered before implementation. The installation of on-line sensors should be carefully selected to facilitate maintenance and recalibration. Many instruments eventually fail simply because they cannot be maintained or recalibrated without shutting down a part of the plant to remove the instrument.

Finally, the calculating power of the computer can be exploited to aid in verifying measurements obtained from on-line sensors. As discussed, solids concentration measurements from sensors were checked by comparing the measured values to predictions developed on the basis of a mass balance. Similarly, process modelling offers the potential for checking the conformance of instrumentation and process performance indicators.

Acknowledgements

This work was financially supported by Environment Canada and the Office of Energy Research and Development Energy, Mines, and Resources Canada. Special acknowledgement is extended to the Ontario Ministry of the Environment and the Tillsonburg WWTP staff for their cooperation and the use of their facilities. Other personnel who contributed significantly to this work included: J. Stephenson, Canviro Consultants, Ltd., and D. Chapman, E. Luxon, J. Matthews, D. Rupke, G. Ward, R. Gillespie, M. Yendt, Wastewater Technology Centre.

Robert D. Hill is control systems engineer at the Metropolitan Denver Sewage Disposal District No. 1, Denver, Colorado. At the time of this study, R. D. Hill was a Visiting Assistant Professor with the Department of Civil Engineering and Engineering Mechanics at McMaster University, Hamilton, Ontario, Canada.

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OPERATIONAL AUDIT ASSISTS PLANT SELECTION FOR DEMONSTRATION OF ENERGY SAVINGS AND IMPROVED CONTROL

by

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Abstract

A real-time microcomputer-based operational audit was used to help select an activated sludge plant for a full-scale demonstration of the benefits achievable through the use of on-line instrumentation and automatic control. Information on aeration system performance and plant hydraulics was obtained to assess the suitability of a candidate plant. The dynamic data obtained indicated that significant savings could result with the implementation of continuous dissolved oxygen control. The candidate plant was selected for the planned full-scale demonstration project.

Keywords

Wastewater treatment; process control; plant operations; aeration; hydraulics; instrumentation; computer; monitoring.

Introduction

In 1983, Environment Canada's Wastewater Technology Centre (WTC) initiated a study to select a full-scale activated sludge plant at which to demonstrate the application of automated process control for improved plant operation. One of the prime goals of the study would be the demonstration of energy savings. This required that significant effort be directed at the aeration process and dissolved oxygen (DO) control since this is generally the largest energy consuming process in wastewater treatment. Very few plants in Canada have attempted automatic aeration control or demonstrated the savings to be achieved or the design tools required to implement control. With an existing plant, there is an opportunity to evaluate or audit actual plant operating performance to ascertain whether there is any advantage to be gained by the application of feedback control. More importantly, a carefully implemented audit could provide a data base upon which to design the electro-mechanical refit of a plant to permit the correct sizing of the process hardware (e.g., motors, blowers, control valves) for the successful application of control. Previous experience had shown that conventional sampling and analysis of a plant provided inadequate information to judge the dynamic performance of a plant or to design a refit for control purposes. Therefore, the WTC decided that an on-site evaluation required the acquisition of dynamic operating data, especially in connection with hydraulic loading and DO concentration to estimate variations in plant loading and the oxygenation capacity of the system. The purpose of this paper is to discuss the techniques used to audit the aeration process at a specified candidate plant, the use of the data to estimate the potential for energy savings through automatic control, and the importance of dynamic data to the plant refit design.

Plant Description

For demonstration purposes, a plant was required having at least two parallel trains--one of which would be automatically controlled and one manually controlled. The

Tillsonburg, Ontario plant was selected as a candidate plant for the study to determine whether it would be suitable for the application of control for energy savings. Although the plant was smaller (9×10^3 $m^3 \cdot d^{-1}$) than originally desired, the hydraulic and air supply configurations appeared to be readily separable into two parallel process trains. The plant had two parallel aeration trains (north and south) with two aeration tanks-in-series in each train. A common 0.30 m diameter air header supplied air to both trains. Two 45-kW and two 30-kW positive displacement blowers were available to deliver air to the common header. This represented apparent flexibility for control purposes if they could be split into equal pairs, each consisting of one 30-kW and one 45-kW blower discharging to separate headers. Each pair would then have been associated with either the north or south aeration train--the north being automatically controlled and the south manually operated for comparison. Separate secondary clarifiers and effluent flow measuring structures were available for each train. Thus, the biological solids inventory could be controlled independently in each train. Each aeration/secondary clarifier train received identical feed since the flow was split at the entrance to the aeration trains.

Materials and Methods

Microcomputer and Data Acquisition Algorithms

An Apple II plus microcomputer with real-time clock and analog-to-digital capability was used to perform on-line data acquisition. Peripherals included a thermal printer, plotter, monochrome monitor and two 5-1/4" floppy disk drives. The data acquisition algorithm which was developed at the Danish Water Quality Institute (VKI) and the WTC was written in Pascal. On-line instrument signals were sampled every ten seconds and the data displayed on the monitor. At six-minute intervals the data were logged to a floppy disk. The stored data were plotted daily. Overall oxygen mass transfer coefficients were determined on-line using a BASIC algorithm. These were measured at 30 kW and 75 kW blower configurations using a non-steady state reaeration technique (Howell et al., 1984) to analyze DO data collected at 90-second intervals after a known step change in blower power.

Dissolved Oxygen Measurement

Laboratory YSI dissolved oxygen instrumentation was used to monitor the D.O. concentration in each of the four aeration tanks. The DO probes were placed at mid-depth, one-third down the length of each first pass and two-thirds down the length of each second pass. Output signal conditioning, isolation and amplification were accomplished at each YSI transmitter using separate 4-20 mA output signal converter/isolation transmitters. Careful consideration of installation, grounding and power supply was required to ensure good overall performance. The probes were calibrated at approximately weekly intervals. Instrument adjustment greater than $0.3\ mg \cdot L^{-1}$ was rare.

Level and Flow Measurement

The secondary clarifier effluent from the north and south trains discharged through a Parshall flume and a rectangular weir, respectively. Capacitance probes were used to measure the liquid level in each structure. The microcomputer calculated and recorded volumetric flow rates. A third level detector was installed at the grit chamber to monitor the input hydraulic disturbances created by the upstream pump station.

Results and Discussion

The hydraulic measurements indicated that the north train received about 20% more flow than the south train over a 24-hour period (Figure 1). On a shorter time scale, the input flow at the grit tank rapidly oscillated from zero to a maximum value due to the on/off pump station operation (Figure 2). The plant hydraulic characteristics effectively damped this rapid input oscillation before it reached the effluent measuring structures.

These data identified that flow balancing would be required and the direction which would be required for better pump station design and control for the plant.

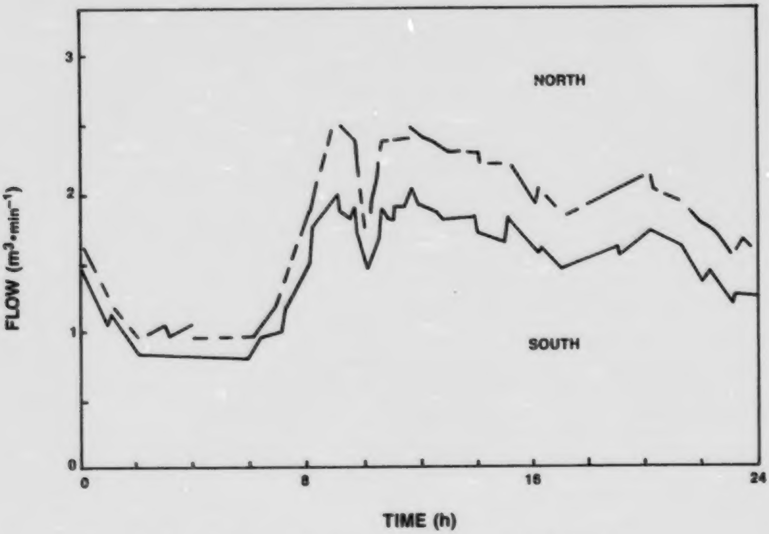

Fig. 1. Diurnal flow measurement in parallel trains.

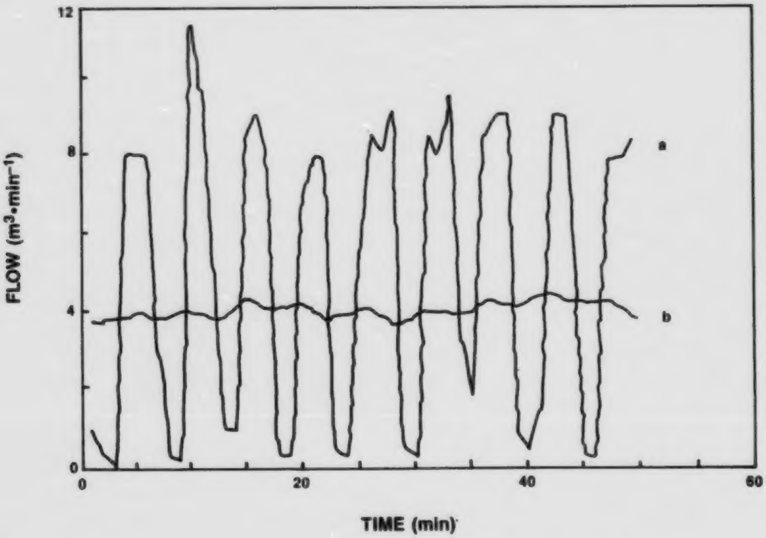

Fig. 2. Influent (a) and effluent (b) hydraulic flow patterns.

The continuous on-line monitoring of the DO in the aeration tanks produced some interesting results. The plant was being operated with 45 kW blower capacity and approximately equal air distribution to the four tanks when the study commenced. The DO concentration was observed to be essentially zero most of the day (Figure 3(a)). The two left hand graphs in Figure 3 represent the first pass of each train while the right hand graphs represent the second pass. With this operating configuration, the load to the two first passes exceeded their oxygenation capacity. Since most of the BOD removal occurred in the first passes, the second passes achieved higher DO's for longer periods, but could not maintain a residual DO except at minimum loadings.

Throughout this time, the visual quality of the plant effluent was excellent. In retrospect, it is believed that nitrification was controlling the DO under these conditions. Plant operating records indicated partial nitrification was occurring, presumably to the extent allowed by the transfer of oxygen into the system, i.e., nitrification was oxygen-limited at peak organic loadings.

Positive DO measurements were required at all times to permit the necessary audit data to be collected. Therefore, the blower output was increased to the 75-kW configuration by starting one of the 30-kW blowers. This immediately resulted in an increased DO in all of the tanks during the off-peak loading period. The DO concentration was higher for a longer period of time in the second passes than in the first passes. However, there was zero DO in the tanks for significant periods of the day even under the 75-kW configuration (Figure 3(b)). The 30-kW blower was shut off after it became apparent that the increased blower capacity alone did not achieve the desired DO profile (at hour 12).

After the 75-kW trials, the airflow was redistributed since it was recognized that the second passes needed less air than the first passes. The airflow to the second passes was reduced to sustain mixing only and force more air to the first passes. With 75-kW blower capacity in operation this achieved the results shown in Figure 4. Although highly dynamic variations in DO are apparent, zero DO concentrations did not occur. Nitrification was known to be complete under these conditions.

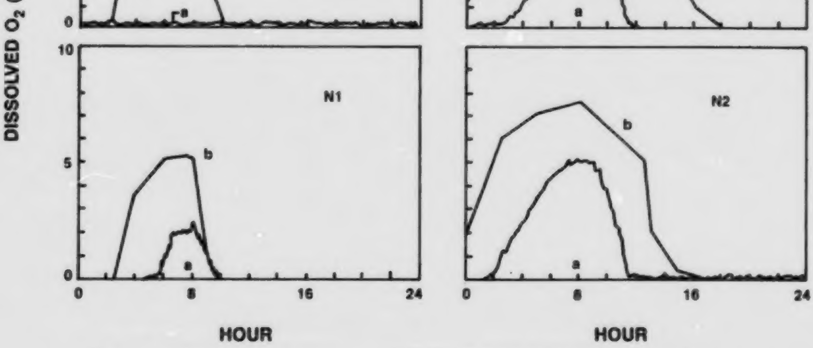

Fig. 3. Dissolved oxygen concentration at 45-kW(a) and 75-kW(b) blower configuration and original air distribution.

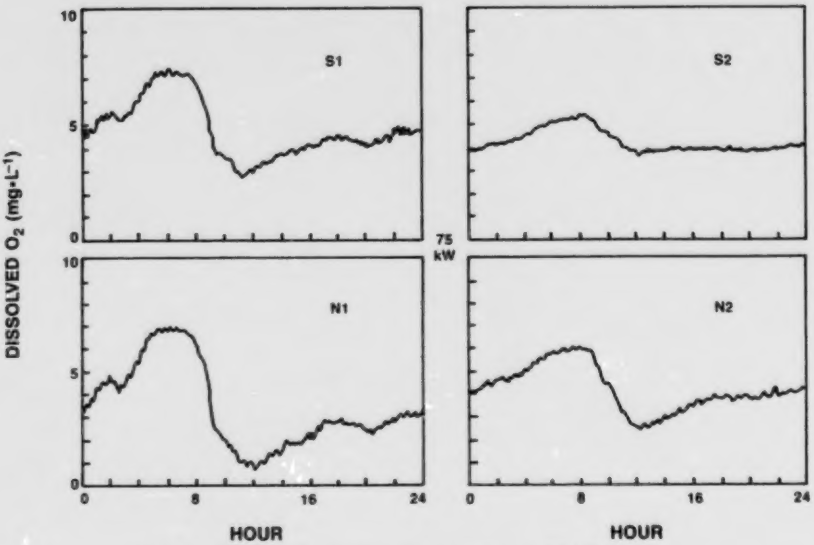

Fig. 4. Dissolved oxygen concentration at 75-kW blower configuration and redistributed airflow.

A measure of the overall oxygen mass transfer coefficient, αK_La, was made at total blower power configurations of 30 kW and 75 kW (Table 1). The air distribution was the same as that which achieved the results shown in Figure 4.

Table 1. Overall Oxygen Mass Transfer Coefficient, αK_La

Total Blower Configuration (kW)	Aeration Tank			
	South 1 (h^{-1})	North 1 (h^{-1})	South 2 (h^{-1})	North 2 (h^{-1})
30	1.15	1.38	1.48	0.44
30	1.39	1.17	-	0.64
75	3.18	2.81	1.68	1.25
75	4.22	3.26	1.48	0.89
75	4.11	4.54	1.40	2.24

For each tank, it was assumed that the relationship between αK_La and the total blower power configuration was linear and passed through the origin. A least squares analysis was used to define this relationship. For example, in tank South 1:

$$\alpha K_La = 0.05030 \text{ (blower power)} \quad (1)$$

The family of curves generated for the four aeration tanks is shown in Figure 5. The ratio of the αK_La estimated for each tank to the sum of the αK_La's for all four tanks at a given blower configuration was used to estimate the airflow being supplied to each tank under process conditions. For example, at the 75-kW blower configuration, the proportion of total airflow supplied to aeration tank South 1 was estimated to be:

$$\% \text{ airflow to S1} = \frac{\alpha K_L a \text{ (South 1)} \times 100}{\sum \alpha K_L a\text{'s (four tanks)}} = \frac{3.77 \text{ h}^{-1} \times 100}{10.23 \text{ h}^{-1}} = 36.9\% \qquad (2)$$

Similarly, the airflows to tanks N_1, S_2, and N_2 were estimated to be 34.2%, 14.8% and 14.2%, respectively. The total airflow for a given blower configuration was determined from the factory blower performance curves.

From DO data similar to those shown in Figure 4 and knowledge of the relationship between the overall oxygen mass transfer coefficients and blower power (Figure 5), the power consumption under simulated DO control conditions was estimated. This required a mass balance on the mixed liquor DO:

$$dC/dt = \alpha K_L a \, (C_s - C) - R \qquad (3)$$

where C = dissolved oxygen (DO) concentration ($mg \cdot L^{-1}$)
C_s = saturation value of DO at process conditions ($mg \cdot L^{-1}$)
$\alpha K_L a$ = overall oxygen mass transfer coefficient (h^{-1})
R = oxygen utilization rate (OUR) or respiration ($mg \cdot L^{-1} \cdot h^{-1}$)

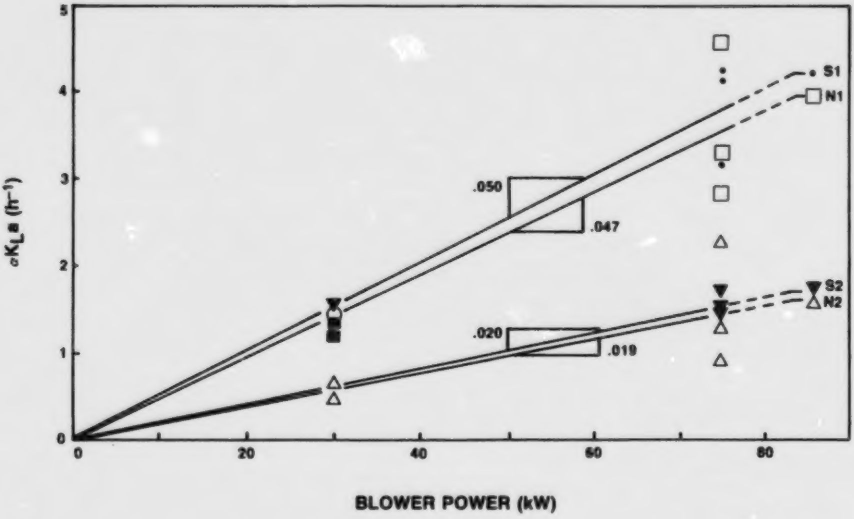

Fig. 5. Relationship of overall oxygen mass transfer coefficient ($\alpha K_L a$) and blower power configuration.

The respiration rate, R, could be calculated from Equation 3 since all the other parameters were known for the conditions under which the on-line DO data were collected. However, it was necessary to recognize that the DO concentration was never at steady state and that the derivative term in Equation 3 was not negligible. Therefore, Equation 3 was discretized as follows:

$$(C_t - C_{t-1})/\Delta t = \alpha K_L a_{t-1} \, (C_s - C_{t-1}) - R_{t-1} \qquad (4)$$

where Δt = sample time interval
t = current sample time
$t-1$ = previous sample time

The respiration rate was calculated by rearranging Equation 4 to:

$$R_{t-1} \quad \alpha K_L a_{t-1} (C_s - C_{t-1}) - (C_t - C_{t-1})/\Delta t \tag{5}$$

Once R was estimated for the actual operating conditions, the total blower power that would be required to maintain the dissolved oxygen concentration at a setpoint level could be simulated. With setpoint control, the DO concentration is constant with time and Equation 5 can be rearranged to give:

$$\alpha K_L a_{t-1} = \frac{R_{t-1}}{(C_s - C_{t-1})} \tag{6}$$

Equation 6 was used to calculate the $\alpha K_L a$ values required to maintain the setpoint DO concentration under variable loading conditions for each tank. From the calculated $\alpha K_L a$ values, and the relationship between $\alpha K_L a$ and total blower power for each tank shown in Figure 5, the total blower power required to maintain the DO setpoint in the respective tanks was estimated. This had to be done separately for each tank since only the total blower power required to deliver air to all tanks through the common header could be measured.

The portion of blower power attributable to air delivery to each tank could not be measured directly. Therefore, the power required for setpoint DO control in each tank was estimated as if only that tank, with its variable loading and fixed air distribution valve position, were controlling the total blower configuration. Figure 6 presents examples of the power consumed by the whole system for the four cases in which the DO level in one tank is controlled at 1.0 mg·L^{-1}. Only the tank used to control the total blower output would maintain its dissolved oxygen concentration at the setpoint value. The DO levels in the other tanks would vary from their respective setpoint values if the

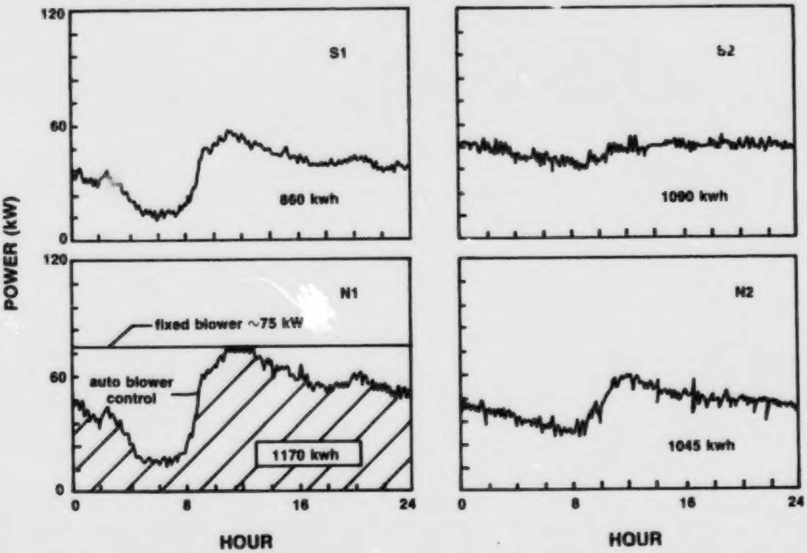

Fig. 6. Total system blower power configuration for simulated 1.0 mg·L^{-1} setpoint DO control (separate case for each tank).

total blower power requirement determined by the controlling tank was different from that calculated to be required for setpoint maintenance in the other tanks. For example, to maintain a DO of 1.0 mg·L^{-1} in tank North 1, the total system power draw was calculated to be 1170 kWh by integrating the area under the power curve for one day of operation. Without control (i.e., constant blower output), the power requirement to maintain at least 1.0 mg·L^{-1} DO in tank North 1 would have been 75 kW--resulting in a 24-h energy consumption of 1800 kWh. The reduction through automatic control (1800-1170 kWh) represents an approximately 30% energy saving over that consumed by fixed blower operation. This is likely a conservative estimate because air distribution control would also minimize the air header pressure while controlling the DO in all tanks, further reducing energy consumption for the total system.

Since this analysis indicated a strong possibility of demonstrating substantial energy savings using aeration control at Tillsonburg, the plant was selected for the purpose of demonstrating at full-scale the use of on-line instrumentation and control. Although the details of the refit design are beyond the scope of this paper, the computer-based, on-line audit provided much of the data upon which the design was based, e.g., required variable speed range for the blowers, and air distribution valve coefficients (C_v). Other factors such as improved biological solids inventory control through automation will also be investigated to determine the benefits in relation to improved plant operation and energy savings.

Summary and Conclusions

The Wastewater Technology Centre demonstrated the use of a portable, microcomputer-based, real-time monitoring system to assess and modify the operational performance of an existing full-scale activated sludge plant.

The real-time dynamic operating data obtained using the audit equipment were valuable in determining the existing capacity of the aeration equipment and the present operating conditions at the plant. This type of continuous data is new to plant operators. It is especially useful and important to regulators and to plant owners and operators who are anticipating plant refits or expansions. A plant expansion or refit could be deferred if such authoritative data indicated adequate plant capacity was available with or without readily modified operating practices. It can provide the necessary dynamic information for correct specification of process hardware such as blowers, control valves and diffusers, to provide operational flexibility and controllability under dynamic operating conditions when plant modifications are necessary.

A study of the plant aeration energy requirements demonstrated that to maintain a minimum DO concentration of 1.0 mg·L^{-1}, an estimated 30% aeration energy could be saved if the plant was refitted for aeration control, compared to fixed blower operation.

The on-line operational audit is a valuable tool which can provide much more process information, with less effort, than can be obtained through manual or discrete sampling.

Acknowledgements

This work was financially supported by Environment Canada and the Office of Energy Research and Development, Energy, Mines and Resources Canada. Special acknowledgement is extended to the Ontario Ministry of Environment and the Tillsonburg WPCP staff in particular for use of their facilities. Wastewater Technology Centre personnel who contributed significantly to this work included: E. Luxon, J. Matthews, G. Horn, M. Yendt, and R. Gillespie. At the time of the study J. Stephenson was employed with the Wastewater Technology Centre.

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CONTROL OF POLYMER ADDITION FOR SLUDGE CONDITIONING: A DEMONSTRATION STUDY

by

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Abstract

The need, for automatic control of sludge conditioning systems, resulted in the development of the sludge conditioning controller (SCC). The fundamental concept behind the SCC and the development of the commercial prototype are described in the paper. The final phase of the program involved demonstrating the operation of the SCC at a full-scale dewatering plant. The SCC was installed on a two metre belt press, dewatering anaerobically digested sludge. The evaluation was conducted by comparing the results of historical manual operation with those achieved on the same press under automatic SCC control. Analysis of the data indicated that an average 24% polymer saving could be achieved by the automated system. The paper also describes the practical problems that were encountered during the installation of the control system.

Keywords

sludge; rheology; conditioning; dewatering; automatic control; instrumentation; costs; demonstration.

Introduction

Sludge handling and disposal represents a major cost component of any sewage treatment system. Sludge volume reduction or dewatering can be instrumental in reducing this cost. With few exceptions, a dewatering system must be preceded by some form of conditioning in order to flocculate the sludge and enhance the ease with which water may be removed. Chemical conditioning of sludge represents a significant operation and maintenance cost, and directly affects the quality of the sludge cake produced by the dewatering device. It is therefore important, from both cost and performance considerations, to optimize the addition of chemicals for sludge conditioning. A variety of tests including capillary suction time (CST), specific resistance, solids concentration, etc., are available to measure the effectiveness of a particular conditioning option. Unfortunately, none of these test procedures have been adapted to on-line measurement. The result is that, as the characteristics of the incoming sludge change, there is no method of automatically adjusting the flow rate of conditioning chemical to compensate for these changes.

The key to increased control of the conditioning process lies in a better understanding of fundamental intrinsic sludge properties such as rheology. Rheological measurements are sensitive to changes in the physical properties of sludge, and existing equipment can be adapted for on-line measurement of rheological properties. Process research conducted at the Wastewater Technology Centre (WTC) has shown that direct measurement of the rheological properties of sludges can be used to control chemical conditioning. This work has been developed and commercialized in the form of a sludge conditioning controller by Zenon Environmental Inc. The paper discusses the rationale behind the control strategy, describes the commercial controller and presents results from the demonstration study.

Background

Rheology can be defined as the study of the properties and behaviour of matter in the fluid state. Most single-phase fluids exhibit Newtonian behaviour in that the rate of viscous flow is proportional to the shear stress.velopment of the commercial prototype are described in th The addition of solid particles to the fluid interferes with the free flow of the dispersion medium to a degree that is dependent on the rate of shear. Sewage sludge behaves in this manner and is referred to as a dispersed system.

The rheology of sewage sludge is further complicated by the fact that most sludges are also thixotropic, meaning they possess an internal structure which breaks down as a function of time and shear rate. A flow curve or rheogram (Figure 1) of a typical sludge, showing the curves produced by a rotating viscometer during the increasing and decreasing rate of shear cycles, indicates that the rheology of the sludge has been altered during the initial (increasing) phase of the test. The displacement of the two curves is referred to as a hysteresis loop and is a measure of the degree of thixotropy exhibited by the sludge.

The relationship between sludge rheology and polymer addition was established during a sludge characterization study at the WTC (Campbell et al., 1978). Figure 2 shows a typical set of rheograms for a polymer conditioned anaerobically digested sludge. As the polymer dosage increased, the shear stress also increased for any given shear rate. At some polymer dose, which was specific to the sludge, a peak appeared in the initial portion of the rheogram. It was observed that the polymer dosage at which this phenomenon occurred, correlated well with the dosage which produced efficient performance during concurrent pilot-scale dewatering trials. These observations formed the basis for selecting rheological properties as a means of automatically controlling polymer conditioning. This assumes that rheograms describe fundamental sludge properties. As the amount of polymer in a system is increased, the nature of the sludge particles or flocs change and subsequently the interrelationship between the dispersed particles (sludge flocs) and the fluid also changes. This, in turn, is reflected in changes in the characteristics of the flow curves or rheograms as shown in Figure 2.

The development of the sludge conditioning controller has evolved through a number of stages. The background work conducted by the WTC included a series of bench-scale batch experiments, operation of a pilot-scale continuous flow conditioning system and testing of an experimental control system on a 0.5 m belt filter press. The objectives of these three phases were to determine the sensitivity of rheological measurements to changes in sludge characteristics and to develop and test an algorithm capable of controlling polymer addition. Details of this work are available in the literature (Campbell and Crescuolo, 1982, 1983, 1985). The rationale on which the control algorithm is based is illustrated by the flow diagram in Figure 3. A reference rheogram is developed and stored in the computer memory. During operation a new rheogram is produced and its characteristics compared with those of the one stored in the memory. Based on this comparison, the algorithm determines what action should be taken by the polymer pump. The entire process is then repeated at pre-selected time intervals.

At the conclusion of phase three, it had been demonstrated that the control strategy developed by the WTC was successful and patents were filed by Canadian Patents and Development Ltd. on behalf of the Canadian Government. The main drawback of the initial experimental system was that the viscometer and computer were too sophisticated and expensive for routine use at a treatment plant. While the rationale behind the control strategy had been shown to be valid, the practical application of the technology depended on the development of an instrument which was not only rugged enough for installation at a sludge dewatering site but was also sufficiently inexpensive to be economically attractive. At this point the WTC entered into a co-operative program with Zenon Environmental Inc. Under the Program for Industry/Laboratory Projects (PILP), Zenon received financial support to develop a commercial prototype

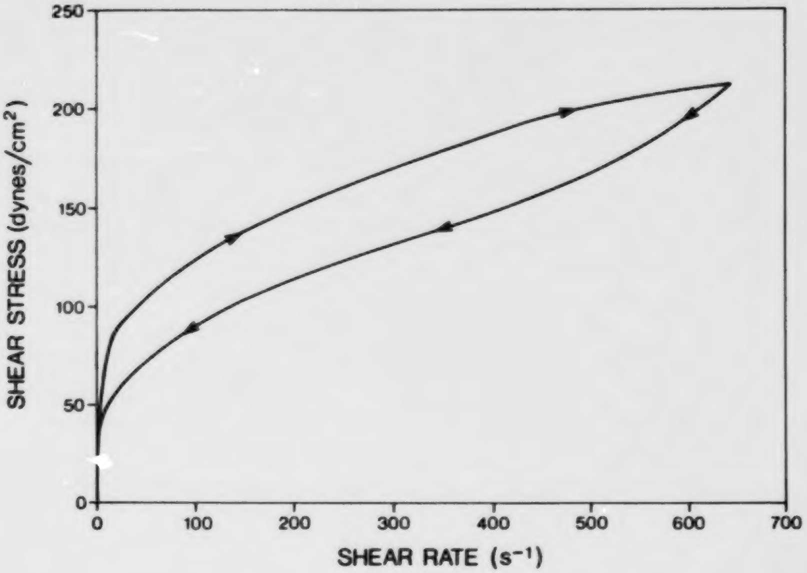

Fig. 1 Typical Rheogram Illustrating Thixotrophy

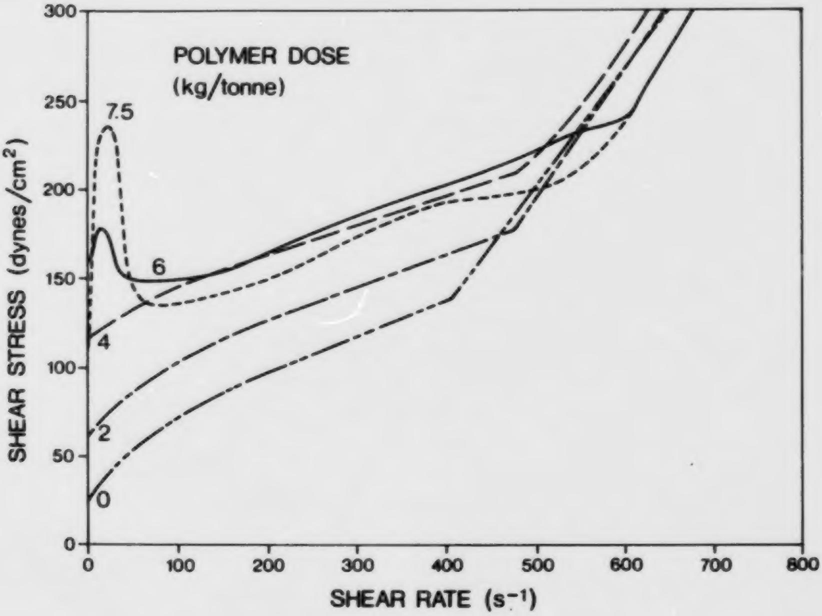

Fig. 2 Rheograms of Polymer-Conditioned Sludge

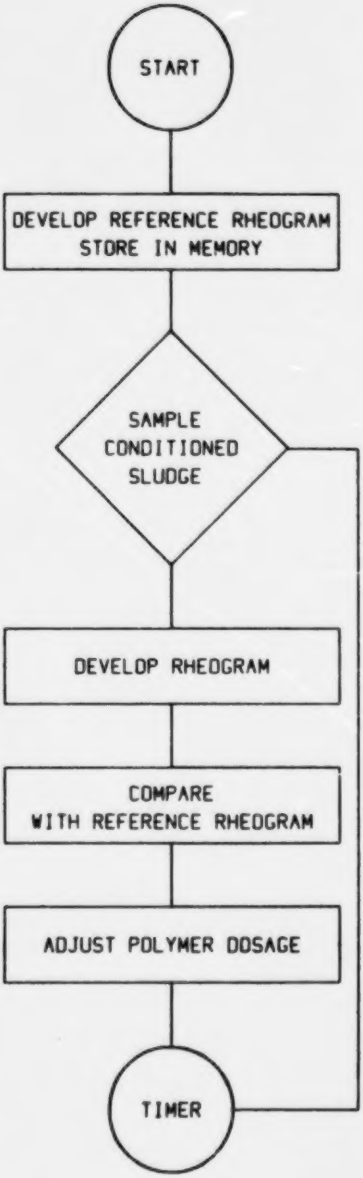

Fig. 3 Flow Chart of Control Algorithm

controller based on the WTC research. The development of the controller was carried out by introducing Zenon designed components to the original WTC test facility one at a time. This enabled each element of the system to be tested independently of the other components. Details of the development and testing of the Zenon sludge conditioning controller (SCC) are available in the literature (Campbell et al., 1986).

Commercial Prototype

The SCC, shown in Figure 4, consists of two major subsystems: the sampling system and the central control panel. One sampling vessel is required for each independent polymer dosing unit while the central control panel is configured to handle up to four sampling systems.

The sampling system consists of a ten-litre tank, the feed and drain pinch valves (5 cm diameter, air operated), three spray rinsing nozzles and the measurement head assembly. The measurement head includes the rheology sensor and its drive motor (sensor head), a signal conditioning amplifier and a level sensor. A local control station is provided near the vessel which allows for manual operation of the feed, drain and spray solenoid valves.

The sample piping is designed to fill the vessel from the bottom. This minimizes the shear to the conditioned sludge. The pinch valves provide an unobstructed flow path when open and positive shut-off when closed. This arrangement effectively eliminates clogging of the sample piping. The vessel itself is constructed of stainless steel. It has a conical bottom to facilitate rinsing. The three rinse nozzles are arranged around the vessel at the level of the rheology sensor. The fan spray pattern from each nozzle provides excellent cleaning of the probe and sufficient overspray to rinse the level sensor and the vessel walls. The level shut-off sensor operates on the basis of conductivity. The probe itself is a piece of stainless steel rod. The lack of moving parts and simple design make it very resistant to fouling.

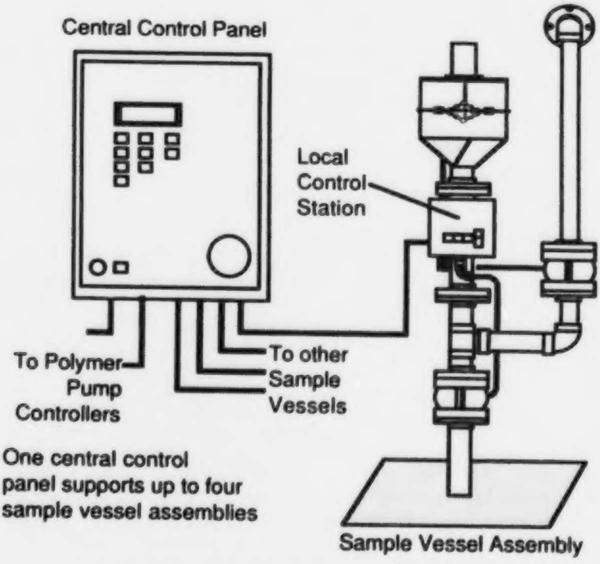
Central Control Panel
Local Control Station
To Polymer Pump Controllers
To other Sample Vessels
One central control panel supports up to four sample vessel assemblies
Sample Vessel Assembly

Fig. 4 Schematic of Sludge Conditioning

The central control panel houses the custom microcomputer and associated hardware. The computer sequences the fill, measurement and drain operations for each sampling vessel. It also receives the raw rheological measurements and calculates the parameters of interest and output control signal for the appropriate polymer pump controller. The operator interface is on the front of the control panel. This includes a series of push buttons to toggle between manual and automatic control modes, increase or decrease the desired setpoints, and check the status of the system and each sampling vessel. System parameters are shown on a two-line LCD display as are any alarm messages.

The system can operate in two modes. The manual mode allows the operator to directly control the polymer dosing pump. In automatic (normal) mode the SCC tracks changes in the dewaterability of the sludge to keep the press operating at optimal conditions. During automatic operation, flocculated sludge is allowed to flow under gravity into the bottom of the sample vessel. The sample is taken from the top of the flocculator or from a header pipe just before the sludge is introduced to the press. In either case, care is required to ensure that the take-off point will provide a representative sample of fully conditioned sludge. Once the vessel fills, the rheology sensor is rotated in the sludge and its output is transmitted to the central control panel. After the measurement cycle is completed, the sample is allowed to drain out and the vessel is thoroughly rinsed for thirty seconds.

Because sludge characteristics vary with time, it may be necessary to redefine the optimal setpoints occasionally. This can be done in one of two ways. When the 'tune' button is pushed the system uses the next five sample results as representative of optimal operation and choses setpoints from this data. This is useful when starting up a press or when changing sludges. It requires the operator to bring the press to optimal running conditions. A setpoint adjustment is also provided which allows the operator to make minor adjustments to the operating setpoints of the press without 'tuning' the press.

The control program incorporates a 'minimum seeking' algorithm. Meeting the requirements is a necessary but not sufficient condition of being at the optimal dose. Thus, even if all of the conditions are met, the polymer dose is decreased by a small preselected increment. At steady state the dose will tend to cycle around the minimum value which meets the setpoint requirements. The computer examines the data from each run to determine possible control actions and also to verify that the sample is acceptable. From the data collected the computer decides if the measurement system is operating properly (i.e., the rheology probe is turning in the sludge or is badly fouled). It also monitors the time to fill the vessel and the initial rate of emptying in order to detect possible plugging conditions. If any of these conditions persist over several samples, the operator is notified of the problem by an alarm.

Demonstration Study

The final phase of the program involved demonstrating the operation of the SCC at a full-scale sludge dewatering facility and determining the actual cost savings. The site selected was the Guelph Water Pollution Control Plant which currently operates two, 2 m Komline Sanderson belt presses on anaerobically digested sludge. Zenon installed an SCC on one of the presses and the WTC monitored the process. Some instrumentation already existed at the plant but additional sensors were also installed. All of the sensors were then connected to a data logging system. The instrumentation package included sludge and polymer flow meters, feed sludge and filter effluent suspended solids probes and an HP data logging system.

The demonstration study was designed such that the SCC could be installed and debugged on one press. This press would be operated for a period of time under manual control and the performance monitored by the data logging system. The system would then be operated under automatic control, the performance monitored and compared with the manual operation. Although side-by-side testing of the two presses was an option, it was decided to use only one press for the study in order to eliminate the inherent differences between two different presses.

Based on preliminary data collection carried out prior to the study, the program was designed on the assumptions that the sludge flow rate would range from 7 to 9 L/s, the feed solids concentration would range from 3 to 6% and a minimum polymer savings of 15% would be necessary for the city to justify purchasing the controller.

Immediately following hookup of the sensors, it became obvious that these assumptions were not valid. The sludge feed pump was an open impeller centrifugal type which produced extreme variations in flow rate. A typical plot of flow rate versus time is shown in Figure 5. The periods of zero flow represent time when flow to the press was stopped while the truck collecting the filter cake was emptied. The flow varied from approximately 3 L/s to 8 L/s with occasional fluctuations to 1 or 10 L/s. This resulted in polymer dosages (Figure 6) ranging generally from 5 to 12 kg/t but sometimes much higher. Initially, attempts were made to compensate for the varying flow by including a flow proportionate component in the algorithm but this was not successful. The problem was finally solved by the installation of a positive displacement sludge feed pump.

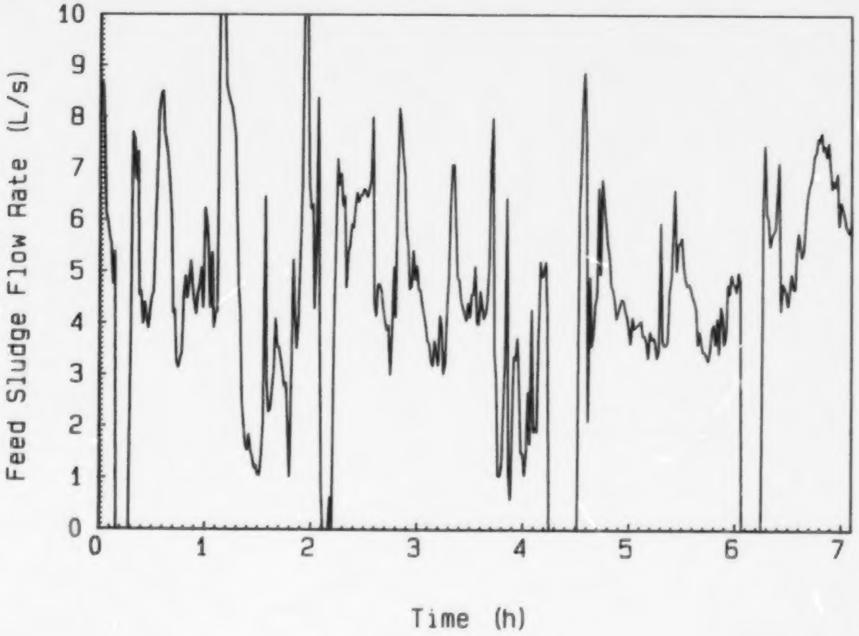

Fig. 5 Variable Feed Sludge Flow Rate

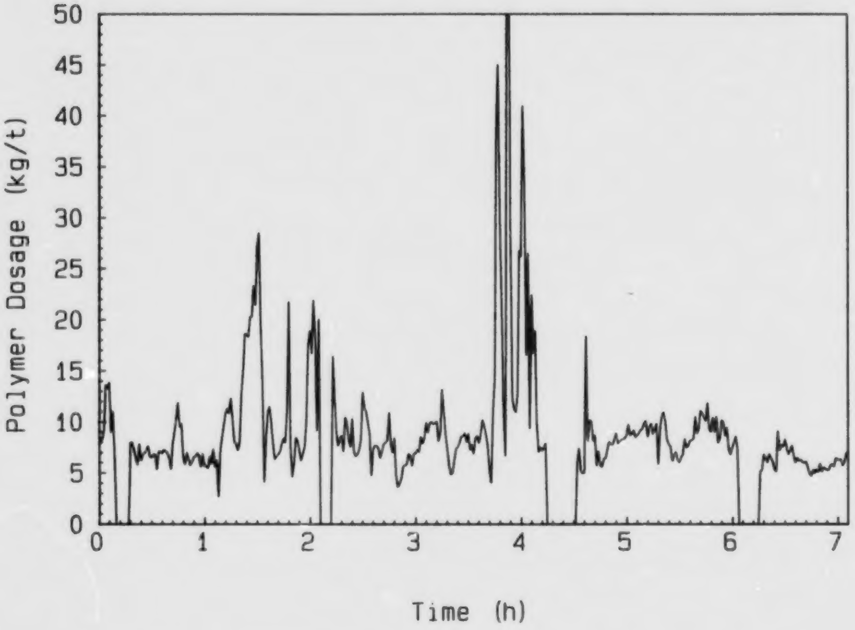

Fig. 6 Polymer Dosage Under Variable Sludge Flow Conditions

The second assumption regarding feed solids concentration was also proven to be false. Figure 7 shows a probability distribution of the feed solids concentration. The median concentration was 2.5%, but 20% of the time the concentration was less than 2%. At extremely low solids levels (i.e., less than 2%), the algorithm could not distinguish between low solids and insufficient polymer addition. Rather than simply specify a minimum operating range of 2% and sound an alarm for the operator when the solids concentration fell below that value, it was decided to increase the sensitivity of the controller down to a solids concentration of approximately 1%. This was accomplished by adding a second sampling port to the system for unconditioned sludge. By conducting rheological measurements on the unconditioned sludge it was possible to determine a relative solids concentration for the sample. This value was then used as a correction factor in the calculations following the measurement of the conditioned sludge. The final operating sequence was modified so that measurements were alternated between unconditioned and conditioned sludges.

With the modifications to the sludge pump and SCC in place, the system operated as expected. Figure 8 shows a test day where the solids concentration ranged from approximately 1.2% to 4%. The sludge feed rate throughout the day was relatively constant at about 6 L/s. In Figure 9, the response of the SCC is presented. As the solids concentration increased, the polymer flow rate increased; when the solids concentration decreased, the polymer flow rate also decreased accordingly. Data from other test days where the solids concentration remained constant, showed that the algorithm was continually searching for a lower polymer flow rate which would satisfy the setpoints. At the completion of the test program the data was analyzed by performing an analysis of variance on the results from the manual and automatic operation periods. Without any adjustment of the data, the degree of scatter was such that there was no statistically significant difference between the two periods. Examination of the complete data set indicated that as the solids concentration decreased, the optimum polymer dosage (kg/t) increased, and that the phenomenon was most pronounced at levels less than 2%. This is shown clearly in Figure 10. The data set from the automatic control phase contained a significantly higher number of days with solids levels less than 2% than did the data set from the manual operation and thus skewed the polymer dosage for the automatic data set towards higher values. In order to compensate for this, data at less than 2% solids were deleted from both sets and the analysis repeated. Under these conditions the automatic operation showed a polymer saving of 12.5% over the manual operation.

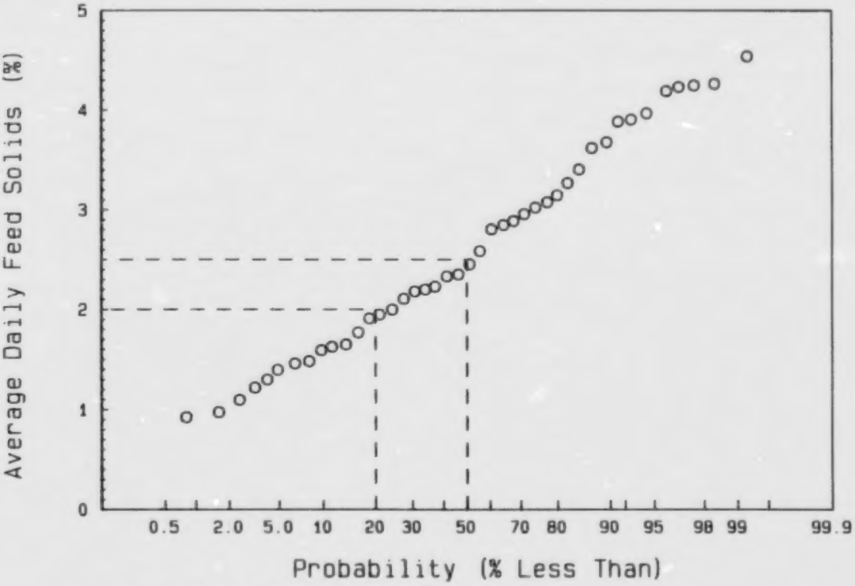

Fig. 7 Probability Distribution of Feed Solids Concentration

However, because it was still believed that the results had been more positive than the above analysis indicated, the entire program was re-evaluated. It had been originally expected that the manual operation period could be directly compared with the automatic period. On closer examination this proved not to be the case. The program had actually consisted of three distinct phases. The first phase was before the constant flow pump had been installed. During this period, control of the polymer flow rate had been minimal. Essentially, the polymer flow rate was set at the beginning of the day and unless some major visible decrease in the performance of the press was noted, the flow rate was not changed throughout the day. Figure 11 shows the polymer flow rate for a typical historical operating day. In the second phase, "manual" operation, the polymer flow rate was adjusted by the operator based on the information that was available to him from the instrumentation, i.e., computer readouts including polymer flow rate, polymer dosage, etc. In the third phase the polymer flow rate was controlled by the SCC. Figure 12 shows probability plots of polymer flow rates on an hourly basis. Comparison of phase 2 (manual with instrumentation) and phase 3 (automatic) show that both phases have approximately the same distribution of polymer flow rates based on hourly averages. The distribution for phase 1 (manual without instrumentation) is

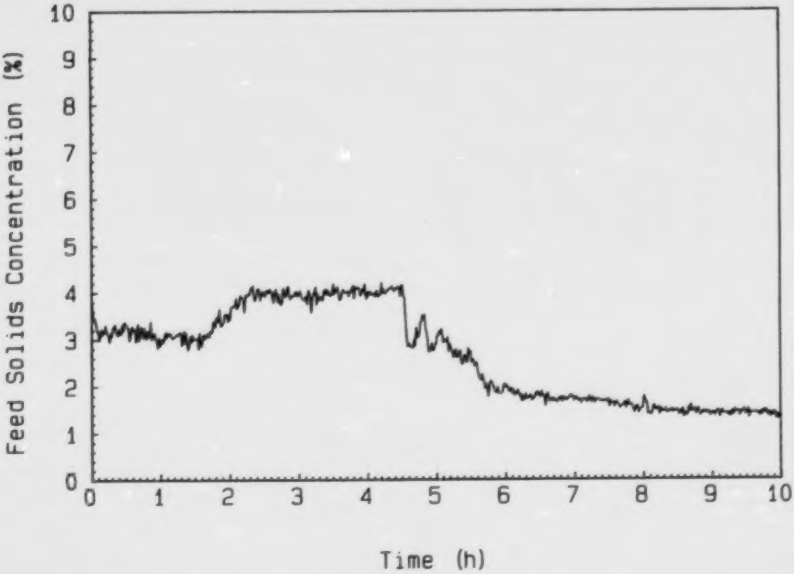

Fig. 8 Feed Solids Concentration Versus Time

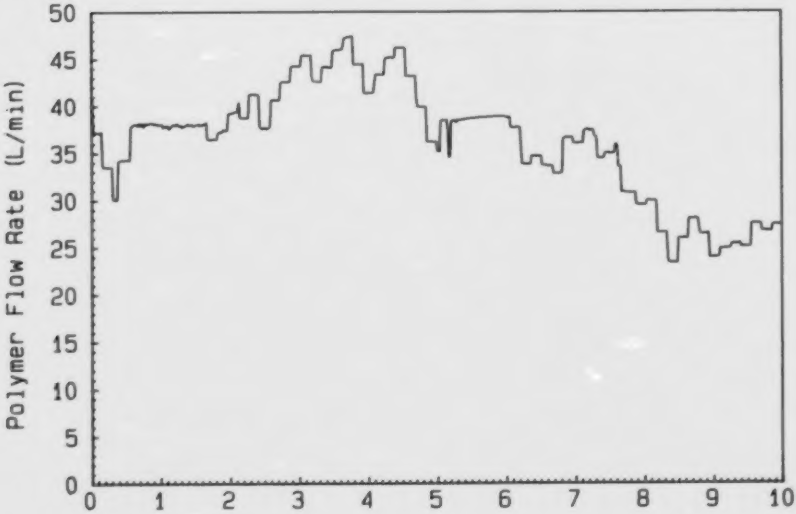

Fig. 9 Response of SCC to Variable Solids Concentration

distinctly different. Instead of being relatively evenly distributed over a wide range of flow rates, approximately 93% of the data falls in the range of 36 to 48 L/min. This compares to a range of 21 to 48 L/min for the second phase and 24 to 57 L/min for the third phase. Thus, the second phase (manual) does not represent the typical manual mode of operation but rather manual operation utilizing a high degree of instrumentation. In order to evaluate the benefits that could be derived from the SCC it was necessary to compare phase 1 with phase 3. Although the first phase with variable sludge feed rate should not be considered as acceptable practice in terms of polymer dosage, it does represent actual current practice in controlling polymer flow rate.

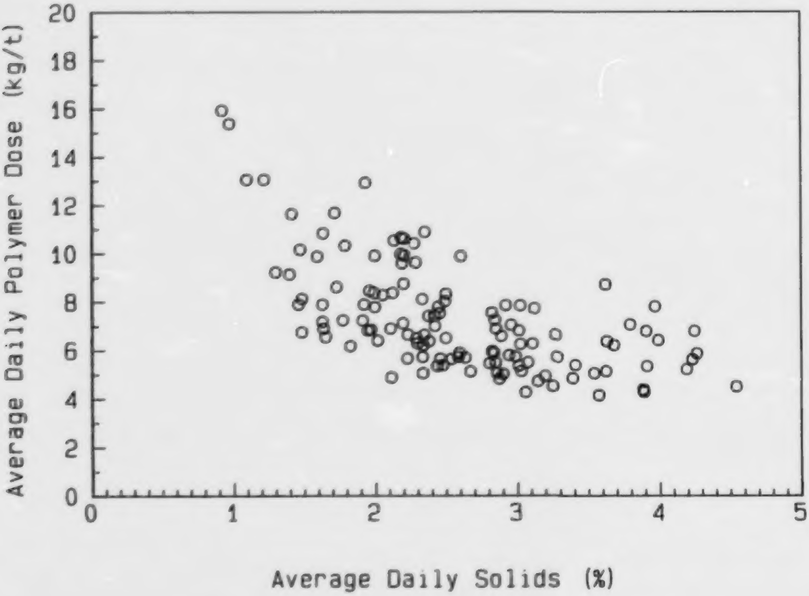

Fig. 10 Effect of Solids Concentration on Polymer Dosage

The method used was to take typical automatic control days and compare the observed polymer flow rate with what would have occurred historically. The median value from Figure 12 (i.e., 40 L/min), was selected to represent the historical polymer flow rate. Figures 13, 14 and 15 illustrate three typical automatic control days. In each case the actual polymer flow rate is shown along with what is assumed to be historical practice i.e., setting the polymer at 40 L/min and leaving it unchanged throughout the day. The area under the two curves was integrated, and the difference between the two areas was assumed to represent the savings due to automatic control. For the test day shown in Figure 13, the sludge solids were relatively consistent at approximately 3% with occasional fluctuations to 2.5% and 4%. The polymer flow rate under automatic control generally fluctuated around the 40 L/min chosen for manual operation, but did have some larger fluctuations which corresponded to changes in the solids concentration. The net effect was that the polymer savings were only marginal at 3.4%. In Figure 14, the solids concentration was very consistent throughout the day at about 2%. The automatic control fluctuated around the optimum dosage, always seeking to find a lower level which would satisfy the setpoints. The polymer savings on this day were 15.9%. For the

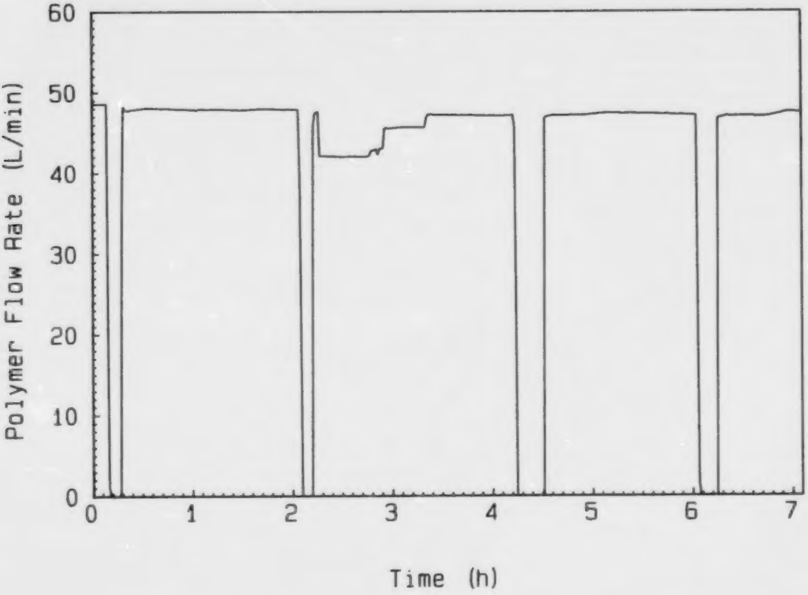

Fig. 11 Polymer Flow Rate Under Historical Manual Control

test day shown in Figure 15, the solids concentration was extremely erratic throughout the day and ranged from 1.4% to 5%. It is this type of situation which best demonstrates the potential of the controller. The controller varied the flow rate from 18 to 40 L/min in response to the changing solids levels. In this case the controller showed a polymer saving of 28.7%.

In order to calculate the average polymer saving due to automatic control, the volume of polymer used for all of the automatic control days after the modifications to the SCC were complete, was determined from the data. The total volume of polymer that would have been used if the pump had been set at the median flow rate of 40 L/min for this entire time was also calculated. Under these conditions the polymer saving was calculated to be 24%. This result, combined with examination of the data, clearly shows that there is a distinct difference between the traditional manual and automatic modes of operation and that significant polymer savings can be achieved.

A number of other aspects were also investigated during the study. Suspended solids in the filter effluent, which was a combination of the filtrate and the belt wash water, was monitored continuously with a solids probe. The results were generally about 0.1% but ranged occasionally from 0.05 to 0.2%. No differences between operating modes were identified but the results of the probe were suspect. During periods of obvious underdosing under manual control, sludge squeezed out from the edges of the belts and fell into the drain area. After mixing with the washwater and filtrate, these solids appeared as extremely large, well flocculated aggregates and there was, based on visual observation, an increase in the solids loss from the system. Examination of the data from the probe never showed any corresponding increase in solids concentration. It is believed that the location of the probe was such that the very large solids resulting from overdosing were not sucked into the probe to be measured. The question of the impact of the controller on solids recovery remains unanswered.

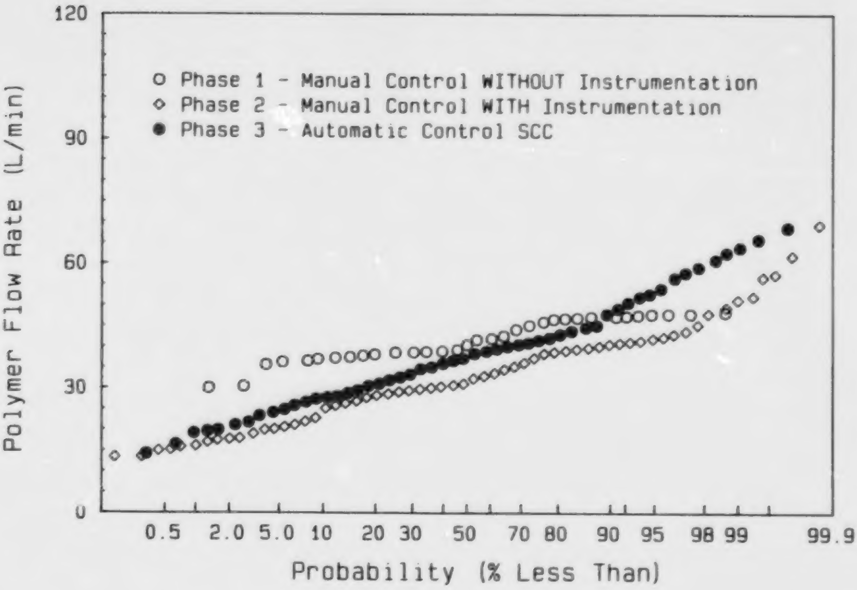

Fig. 12 Probability Distribution of Polymer Rate for Three Test Periods

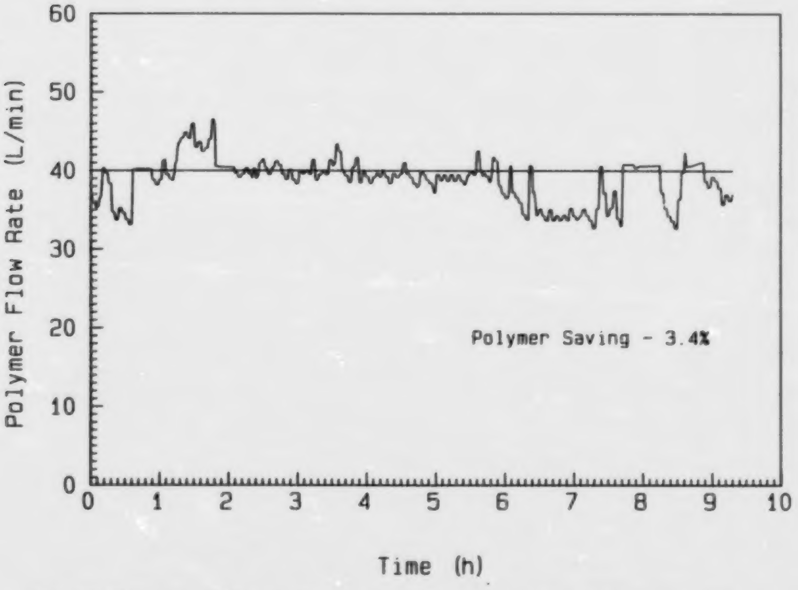

Fig. 13 Response of Controller to Low Variability in Solids Concentration

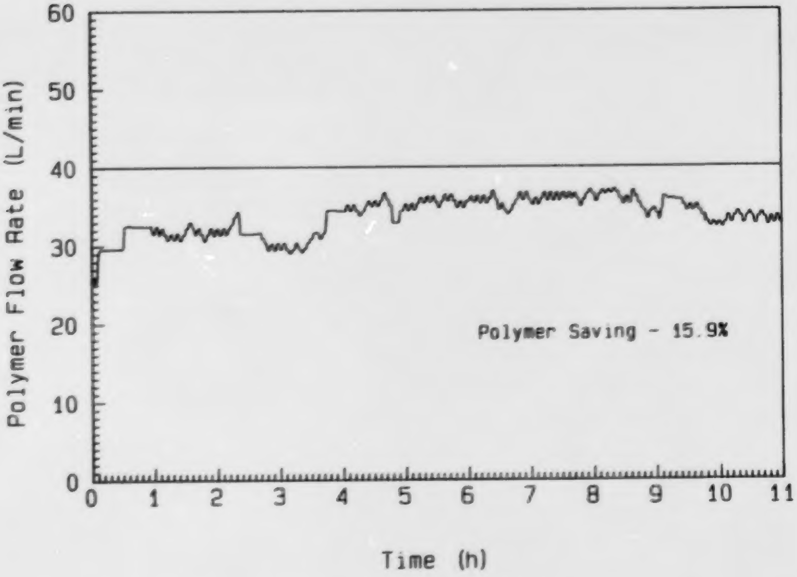

Fig. 14 Response of Controller to Consistent Solids Concentration

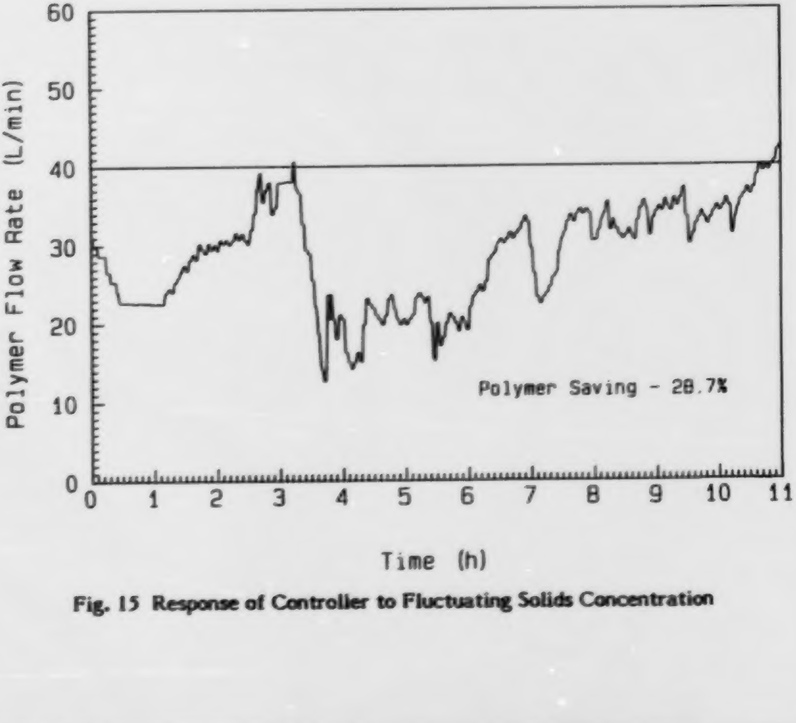

Fig. 15 Response of Controller to Fluctuating Solids Concentration

Cake solids from the belt discharge were sampled by hand and measured on a regular basis. The results were similar to those for the filtrate solids in that no difference between operating modes was identified. The cake solids were relatively low at 17 to 19% and it was felt that with automated control of the polymer dosage, the operation of the belt press could be optimized to improve the cake dryness. This activity was outside the scope of the demonstration program.

A number of improvements to the sludge conditioning controller were implemented during the study. A second sampling port was added to increase the sensitivity of the instrument to low solids concentrations. The spray pattern was also modified to improve the washing of the sensor and the inside of the sampling vessel.

Summary

The sludge conditioning controller can operate without plugging or fouling over an extended period of time. During the entire study, a period of almost seven months, the sensor and sampling vessel were only cleaned manually two times.

Operation of the SCC results in a polymer flow rate versus time relationship which is distinctly different to that observed during manual operation without extensive instrumentation and data logging capabilities.

The average potential polymer savings with the SCC at the Guelph plant is estimated to be 24%, as compared to historical data.

Similar polymer savings can theoretically be achieved by a skilled operator who has a highly instrumented system complete with a data logger. The disadvantage with this approach is that the majority of the operators time will be spent making adjustments to the system, and the cost of an instrumentation package comparable to that used in Guelph is considerably more than the cost of an SCC.

Although confirming data from the study is not available, visual observation of the press throughout the study indicated that solids recovery was improved during periods when the SCC was operational.

Zenon Environmental Inc. are actively marketing the SCC under licence to Canadian Patents and Development Limited.

Acknowledgements

The authors would like to acknowledge the support and co-operation of the staff at the Guelph Water Pollution Control Plant, in particular, Mr. Don Watson, the belt press operator and Mr. John Sanvido, Superintendent Water Department/Water Pollution Control Plant.

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CONVERSION OF SLUDGE TO OIL: A NOVEL APPROACH TO SLUDGE MANAGEMENT

by

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Abstract

Experiments at the Wastewater Technology Centre have shown that low temperature conversion of sludge to oil represents a promising alternative to sludge incineration. A 1 kg/h bench-scale reactor produced oil yields ranging from a low of 13% for an anaerobically digested sludge to a high of 46% for a mixed raw sludge. Char yields ranged from 40 to 73% at the optimum operating temperatures. Preliminary results from a 40 kg/h pilot plant were similar to those obtained on the bench-scale system with the same sludges. The economics of the conversion process are compared with those for incineration. The impact of the "sludge to oil" technology on the design of sludge management systems is also discussed.

Keywords:

sludge; fuel; thermal conversion; energy; economics.

Introduction

Over the past few years there has been a growing trend towards recognizing the importance of the sludge train in both the design of wastewater treatment plants and the evaluation of capital and operating and maintenance costs. The days of the process sheet which ended with an arrow for the sludge saying simply "to disposal" have finally disappeared. A recent study (Proctor and Redfern, 1987) confirmed the statement often found in the literature or in presentations, that sludge treatment costs are in the order of 50% of the total wastewater treatment costs. It showed that sludge treatment costs at four integrated sludge management systems, all utilizing incineration, ranged from $350 to $1042 (1986 Cdn$) per metric tonne of dry sludge. In view of these high costs, the efficiency of sludge management must be improved, either by optimizing and upgrading existing facilities, or by introducing new technology. Although significant cost savings can be achieved by optimization of existing systems, alternative new technologies offer the greatest potential for improved energy recovery and decreased cost. Processes such as starved air incineration, gasification and liquefaction are in various stages of development and a review of these technologies is presented elsewhere (Bridle and Campbell, 1983). The most promising new technology appears to be low temperature conversion of sludge to fuel products.

Background

The basic concept of low temperature conversion of sewage sludge to produce fuel products has been known for many years (Shibata, 1939). Recently, German researchers have made significant advances in understanding the mechanisms by which sludge is converted to oil (Bayer and Kutubuddin, 1982). They heated dried sludge to 300 - 350°C in an oxygen free environment for about 30 minutes. The researchers postulated that catalysed vapour phase reactions converted the organics to straight chain hydrocarbons, much like those present in crude oil. Analysis of the product confirmed that aliphatic hydrocarbons are produced, in contrast to all other pyrolysis processes, which produce aromatic and cyclic compounds, regardless of the substrate (e.g., sludge, cellulose or refuse). The German researchers demonstrated oil yields ranging from 18-27% and char yields from 50-60%. The oil had a heating value of approximately 39 MJ/kg and the char about 15 MJ/kg.

In 1982, Environment Canada embarked on a major technology development and demonstration program, with the objectives of assessing the viability of the oil from sludge technology, and ultimately demonstrating, in conjunction with the private sector, the technology at full-scale. To date, this program has consisted of preliminary batch tests to confirm the German results, development of a continuous-flow bench-scale reactor, testing of several sludges on the bench-scale system, design and fabrication of a pilot-scale conversion system and currently, testing of the pilot system.

Bench-Scale Studies

The bench-scale reactor system is shown schematically in Figure 1. The system has a capacity of 1 kg of dried sludge per hour and has been used to evaluate process performance and to generate design information.

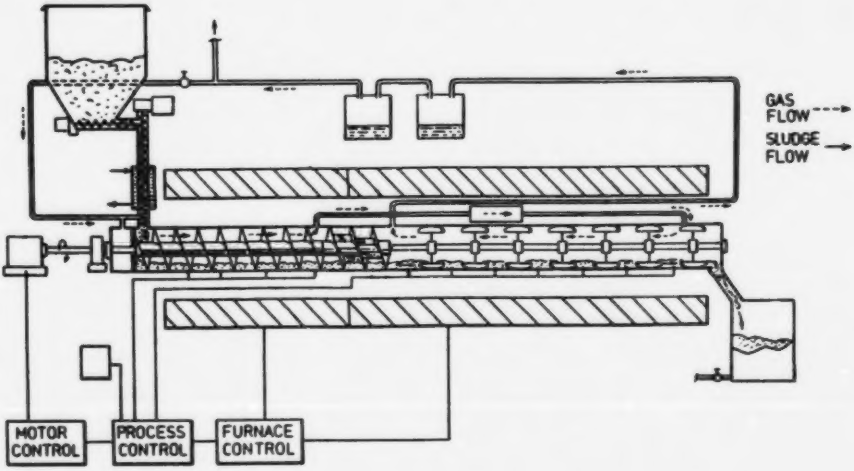

Fig. 1 Schematic of bench-scale reactor

The reactor is subdivided by a helical gas seal into a volatilization zone and a char/gas contact zone. Solids retention time (SRT) in the reactor is controlled by varying both the sludge feed rate and the reactor inventory. Sludge is fed to the reactor by a calibrated screw conveyor and travels through the reactor by means of the reactor conveyor. Volatilized material is withdrawn in the first zone and is contacted with the char in either a co-current or a counter-current mode in the second stage. Product vapours are condensed externally in a water cooled condenser and separated into oil and reaction water. The non-condensable gas (NCG) is vented to a stack. Inert gas is used to purge the system of oxygen and the operating pressure is generally less than 2 000 pascals.

The system has been operated using eighteen different sludges from Canada, USA and United Kingdom. A summary of representative process performance data is shown in Table 1. Additional details on the results achieved during this phase of the development are available in the literature (Bridle and Campbell, 1983, 1984; Campbell and Bridle, 1985).

The information in Table 1 has been separated into mixed raw and anaerobically digested wastewater sludges. It is obvious that the oil yield from raw sludge is generally greater than that from digested sludge. This is as expected because the volatile material which is destroyed through the digestion process represents precursors for the oil which would have been generated in the conversion process. Oil yields range from a low of 13% for an anaerobically digested sludge to a high of 46% for a mixed raw sludge and are primarily a function of the operating temperature and the sludge source. Char yields vary from 40% to 73% at the optimum operating temperature. NCG yields vary from 3% to 12% and are primarily a function of operating temperature. Production of reaction water varies from 3% to 16% and does not appear to be directly related to any of the operating variables. Thermal efficiencies of greater than 95% are routinely obtained with the bench-scale reactor.

From an operational point of view, the reactor has proven to operate equally well with either raw or digested sludge. The main exceptions are the amount of oil produced, (higher quantities from raw sludge) and the oil produced from a specific digested sludge has a lower viscosity than the oil produced from the raw sludge from the same source. In general, oils from digested sludges tend to exhibit lower viscosities than those from raw sludges.

The elemental characteristics of the oil are quite stable over a wide range of operating conditions. Typical values for C, H, O, N and S are 76%, 11%, 6.5%, 4% and 0.5%

Table 1. Representative Operating Conditions and Results

	Raw	Digested
Feedrate (g/h)	750	750
Temperature (oC)	450	450
Solids Retention Time (min)	20	20
OIL		
Yield (%)	24-46	13-29
Viscosity (cstks)	23-62	9-87
Calorific Value (MJ/kg)	33-38	32-42
CHAR		
Yield (%)	40-66	41-73
Calorific Value (MJ/kg)	7-23	6-17
NCG		
Yield (%)	3-10	4-12
Calorific Value (MJ/kg)	2-6	4-6
REACTION WATER		
Yield (%)	3-10	7-16

respectively. Generally the oil contains less than 7% oxygen, but levels as low as 2% have been achieved. Detailed oil characterization studies, conducted under contract by Chemical Engineering Research Consultants Ltd. (CERCL), indicate that specific reactor operating conditions produce an oil essentially composed of saturated alkanes, with some amines still present. An upgrading process currently under development has been successful, at laboratory scale, in removing approximately 50% of the remaining nitrogen and significantly improving the oil/water phase separation.

The effect of temperature on the oil and char yields is illustrated in Figures 2 and 3 respectively. The data for oil yield in Figure 2 has been plotted on the basis of volatile solids rather than total solids. This use of volatile solids as the basis on which oil yield is calculated, in effect normalizes the results to account for the fact that raw sludge generally has higher volatile solids than digested sludge. Even with this normalization, the yield from raw sludge is still generally higher than that from the digested sludge. This indicates that the volatile solids in the raw sludge contain more of the components which are ultimately converted to oil than do those in digested sludge. The volatile solids which are destroyed during digestion constitute a very high percentage of the oil precursors. Although the oil yield varies significantly from sludge to sludge, the effect of temperature is identical for all sludges. The oil yield increases with increasing temperature until it reaches a maximum in the 400 to 450°C range. Above the optimum temperature the yield begins to decrease as conditions favour the formation of increasing quantities of non-condensable gas. Sufficient data is not available to pinpoint the exact shape of the temperature/yield curve but it is obvious that optimum, for all practical purposes, is a range as opposed to one specific point. In engineering terms, this is extremely important because it indicates a very stable operating system (i.e., if the temperature fluctuates to some degree due to other pertubations within the system, the process will not fail). No clear relationship between temperature and oil calorific value has yet been identified but the viscosity of the oil has been found to decrease with increasing temperature, indicating that some thermal cracking of the oil is occurring.

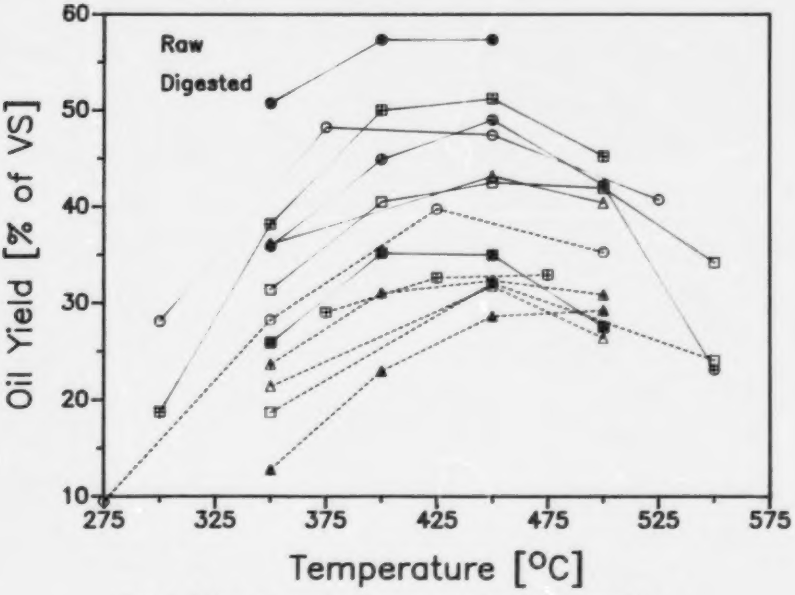

Fig. 2 Effect of temperature on oil yield (bench-scale)

In Figure 3, the char yield has been plotted in terms of total solids because the fixed portion of the solids will ultimately end up as char. In general, the char yield is higher for a digested sludge than for the same sludge prior to digestion (i.e., raw). The relationship between char yield and temperature is very different from that of oil yield. As temperature increases, char yield decreases and continues to do so up to the maximum temperature tested. This is due to the fact that an increasing fraction of the volatile material is vapourized with increasing temperature, leaving less material to end up as char. This also results in the calorific value of the char decreasing as the temperature increases. Since it is generally assumed that the oil is the more valuable and saleable end-product, normal operating practice would dictate setting the temperature to maximize oil yield and accepting whatever char yield occurred.

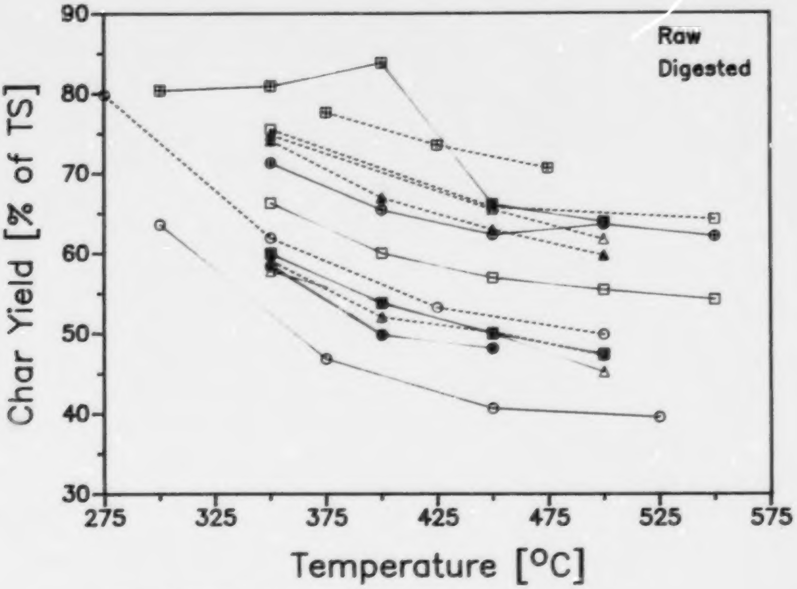

Fig. 3 Effect of temperature on char yield (bench-scale)

The bench-scale program was instrumental in understanding the impact of operating variables on the performance of the system. Data generated during the program was used by Canadian Patents and Development Ltd. to patent the reactor on behalf of the Canadian Government. The information was used to prepare preliminary cost estimates and energy savings for the process, and to scale up from the bench reactor to the pilot plant by a factor of 40 to 1.

Pilot-Scale Studies

In 1985 Environment Canada contracted with Petro Sun International Incorporated (PSI) to design, construct and operate a one tonne per day pilot-scale conversion reactor system to confirm the laboratory-scale results and the projected energy savings from the oil from sludge technology. A schematic of the pilot plant system is shown in Figure 4. It consists of a 1 m^3 sludge storage bin, a sludge feed system, a 25 cm diameter x 3.2 m

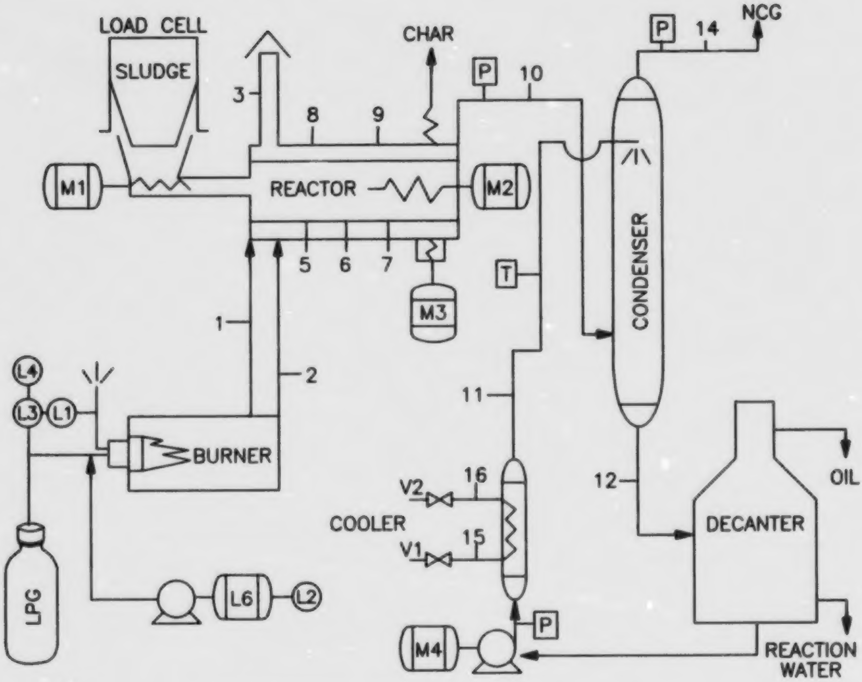

Fig. 4 Schematic of pilot-scale conversion system

long reactor, a char discharge system, a packed-tower direct-contact condenser, an oil-water separator and a self-contained propane/NCG fired boiler to provide hot flue gas for reactor heating. The configuration and operation of the reactor is exactly the same as described for the bench-scale reactor with the exception of the method of heating. The entire pilot plant is mounted on a 1. 7 m x 7.4 m skid and is readily transportable. The system is located at the Regional Municipality of Hamilton-Wentworth, Woodward Avenue Sewage Treatment Plant. It has been debugged and commissioned and is currently undergoing extensive process performance testing.

Some results comparing the performance of the pilot plant with the bench-scale unit for two stabilized sludges, are shown in Table 2. The bench-scale data represents the results from the optimum run on that sludge. The pilot plant data represents the results from a number of runs but because analysis of the products is not complete, some parameters have only single values while some such as NCG have no values as yet. Since the methods for measuring yields are still being refined, the data from the pilot plant must be considered as preliminary at this time. In terms of gross parameters the comparison between the two systems is quite acceptable. The oil yields and the calorific value of the oils are essentially the same. The oil from the pilot plant, for sludge H, appears to have a slightly lower viscosity than that from the bench-scale system. Elemental analysis of the pilot plant oils is in progress. The char yields from bench and pilot-scale runs are generally comparable but calorific values are not yet available. The production of reaction water is similar for both plants.

At this stage of the development program the pilot plant is functioning well from the viewpoint of materials handling and, in general, the characteristics and quantities of end

Table 2. Comparison of Bench-Scale and Pilot-Scale Results

	Sludge R		Sludge H	
	Bench	Pilot	Bench	Pilot
Feedrate (kg/h)	0.75	17	0.75	24
Temperature (^oC)	450	450	425	425
Solids Retention Time (min)	15	20	17	20
OIL				
Yield (%)	15.6	10-14	12.3	9-13
Viscosity (cstks)	14.4	14-20	32.4	21-26
Calorific Value (MJ/kg)	42.5	43-45	41	41
CHAR				
Yield (%)	62.8	52-71	73.4	61-69
Calorific Value (MJ/kg)	8.6	na	5.9	na
NCG				
Yield (%)	8.1	na	3.3	na
Calorific Value (MJ/kg)	4	na	7.0	na
REACTION WATER				
Yield (%)	10.6	10-17	6.6	5-12

na = result not available

products are as expected. The current experimental program will test a minimum of three different sludges and will determine the effect of operating parameters such as temperature, solids retention time, feed rate, condensing temperature and operating mode of the reactor on the performance of the system. Limiting values for feed rate, temperature, etc. will also be determined and this information will be used for the design of a full-scale system capable of processing 20 to 65 tonnes per day.

Economics

Thermal based technologies such as incineration are expensive methods of sludge disposal. When other less complex alternatives for sludge disposal such as land utilization or ocean dumping, are available and acceptable, they are invariably less expensive. However, these alternatives are tending to become less suitable for large urban municipalities for a number of reasons. For example, sludge from heavily industrialized cities may not be acceptable for land utilization due to excessive heavy metal content, and the availability of land within a reasonable transportation distance from large treatment plants is decreasing. Ocean disposal is not an option for inland cities and even where it is the current method of disposal, the time frame in which this practice will be allowed to continue, is very uncertain. It would therefore appear that in the future there will be a growing trend by large municipalities towards some thermal based method of sludge disposal, since it is this type of technology which produces the smallest amount of material (i.e., ash) requiring ultimate disposal. In order for conversion to be accepted in the market place it must be cost competitive with the incineration systems which are currently in operation.

The first step in developing a cost comparison is to ensure that accurate costs are available for existing technologies. This is not the situation for sludge handling facilities. In many cases sludge facilities have been constructed in a number of phases over several years and plant personnel do not have a good estimate of the overall costs. In order to address this problem, Environment Canada contracted with Proctor and Redfern Ltd. to develop total sludge treatment costs at four integrated sludge management plants in Ontario (Proctor and Redfern, 1987). The sludge treatment train was considered to include digestion, thickening, dewatering, conditioning, incineration and ash disposal, although not all unit processes were present in each plant. Capital costs were taken from actual construction contracts and updated to July 1986 by applying the Engineering News Record Construction Cost Index - Toronto. Operating and maintenance costs were determined from plant records and discussions with plant personnel. The costs were presented in a variety of ways but the most useful method was determined to be a total cost (O&M plus amortized capital) per unit tonne of raw sludge generated. The capital component was calculated by amortizing the total capital cost over 20 years at 10% interest. At existing flows, which range from 36 to 67 tonnes per day (on a dry solids basis), the total sludge costs range from $350 to $1042 (1986 Cdn$) per tonne. This represents 35%, 50%, 51% and 52% of the total wastewater treatment costs, for the four plants surveyed. The sludge operating costs represent 48 to 60% of the total plant operating costs while the capital costs related to sludge, account for 32 to 51% of the total plant capital costs.

Projected costs for the oil from sludge technology were developed in two stages. Preliminary costs, based on data from the lab-scale unit, were prepared for a 25 tonne per day plant in 1985 by Zenon Environmental Inc. These costs were updated by Petro Sun in 1986 based on additional data and scaled up to 65 tonnes per day, which compared favourably with the capacity of one of the plants surveyed in the Proctor and Redfern study. This plant had the lowest cost of the four plants surveyed at $350 per tonne. The estimated capital cost for the conversion plant includes dewatering, conditioning, drying, conversion, condensing, oil/water separation, char combustion and ash disposal. It does not include anaerobic digestion because digestion is not a necessary component of a conversion plant. The capital costs are based on adjusted quotes from suppliers for specific unit processes and amortized over 20 years at 10% interest. Operating costs are based on an estimate of the manpower requirements for each unit process. Energy requirements are based on theoretical parameters and data generated from the bench-scale unit. Maintenance costs are based on a percentage of the total equipment cost. The cost of the 'turn-key' plant is estimated at $18.6 million, which is equivalent to a unit cost of $207 per tonne. It was assumed that the oil yield would be 30% on a total solids basis and that the oil could be sold for $0.20 per litre. When the revenue value of the oil is factored into the analysis, this reduces the sludge treatment cost to $148 per tonne as compared to the minimum cost of $350 per tonne reported by Proctor and Redfern.

It should be noted that the costs generated by Proctor and Redfern represent the costs required to build the plant as it existed in 1986. It does not account for the fact that if that plant were being designed today, many improvements and operating strategies, which represent current state-of-the-art, would be incorporated into the design and result in a significantly lower cost. However, the projected cost savings with conversion are of a sufficient magnitude to warrant demonstrating the technology at full-scale and confirming whether these estimates can be achieved in actual practice.

Impact of Technology

Thermal conversion of sludge to fuel offers the wastewater industry a number of potential advantages. The first is the projection of reduced cost for municipal sludge treatment, as discussed in the previous section.

The second is an increase in the energy efficiency of sludge treatment systems, not only in the generation of that energy but also its utilization. In the past, it has been standard practice to anaerobically digest sludge and produce methane as a source of energy. The methane was generally used to heat the digesters and to provide heating for the entire physical plant when required. Unfortunately, in much of North America there is not sufficient methane to satisfy all of the heating requirements in the winter, and there is an excess in the summer. This results in purchasing auxiliary fuel in the winter and flaring excess methane in the summer, unless some other means of using the energy is designed into the plant. Another means of recovering energy is to utilize the hot flue gas to generate steam and ultimately produce electricity. This process is inherently inefficient due to the number of energy transformations which are required, i.e. from thermal to mechanical to electrical energy. Once the electrical power is generated, a market for it must still be found and in Canada at least, local utilities will generally pay only marginal rates for each kwh generated. By comparison, the production of oil by thermal conversion is very energy efficient. The oil is generated in one catalysed thermo-chemical conversion which is much more efficient than transformations between mechanical and electrical energy. In terms of thermal energy, the oil from sludge process is 95 to 98% efficient (i.e., 95 to 98% of the energy in the dried sludge is recovered in the various products). The other major advantage that this process offers is flexibility. The end product can be stored and transported so that the oil can be used on or off site and it can be used only when needed. This factor offers a multitude of ways in which the oil can be utilized such as in-plant use, use within the municipality and use as a feed stock in a refinery.

The third area in which this technology offers a major advantage is in the overall philosophy of treatment plant design and operation. One of the disadvantages of high-rate treatment systems has traditionally been the increased sludge production and its associated downstream treatment costs. With conversion, increased sludge quantities will result in increased oil production, thus the increased cost of sludge treatment will be offset by increased revenue from the oil. Anaerobic digestion has been an integral component of most sludge treatment systems and has been used both to stabilize the sludge and to reduce the volume for subsequent treatment. As discussed earlier, the generation of methane is not a particularly efficient method of recovering energy and all of the volatile material which is converted to methane, is material which cannot be converted to oil. The elimination of digesters represents a significant cost saving in the capital cost of a treatment plant. The dewatering of raw sludge as opposed to digested sludge can also offer advantages in both cost and performance.

The above discussion is based on the assumption that the construction of a complete new sludge treatment facility is required. The conversion process also has the potential to be integrated into an existing plant in a retrofit mode. For example, a portion of the sludge could be converted to oil to provide fuel to incinerate a non-autogenous sludge or an existing incinerator could be used as the char combustor. Although the economic advantages are not expected to be as dramatic as in the case of a 'turn-key' facility, the process may significantly reduce sludge management costs and increase flexibility.

The implementation of the oil from sludge technology into the wastewater industry could be accomplished by private ownership/private operation, public ownership/private operation or public ownership/public operation. The combination of private ownership and operation probably offers the greatest potential for impact on the industry. If a private company has a long term contract with a municipality to dispose of their sludge for a set tipping fee, then the profit generated by that company will be a function of the quantity and quality of oil produced. This will put operation of the plant into the same context as a chemical process plant which produces a saleable end product and where profit is a direct function of process performance and efficiency.

No unique environmental problems are expected to result from implementation of the conversion technology. The reaction water and non-condensable gas can be safely disposed of in the char combustor while at the same time recovering the energy in them. Material balances have shown that essentially all of the metals are retained in the char. These will remain in the ash after combustion, in the same form as in incinerator ash (i.e., essentially immobile) and should present no disposal problems. An intensive program is currently underway at the WTC to evaluate any environmental implications of combusting the oil.

Current Status of Program

The pilot plant has been debugged and commissioned, and preliminary results appear to be comparable to those generated on the bench-scale system. An extensive experimental evaluation of the pilot plant is in progress and should be completed by early 1988. Efforts are currently underway to select a site for a full-scale demonstration plant. It is expected that the design of the full-scale plant will be completed during 1988, with construction scheduled for 1989/90 and start-up in late 1990.

Acknowledgements

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IMPROVING HYDRAULIC EFFICIENCY IN HIGH RATE ANAEROBIC SYSTEMS

by

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Abstract

Maintaining long term organic removal efficiency in a high rate anaerobic process requires the prevention of excess biomass accumulation and its negative impact on reactor hydraulics. However, direct measurement of accumulated biomass has proven to be difficult in the closed, fixed-film reactors which are typical of these systems. Through extensive pilot scale experimentation, a tracer response and regression technique was developed for indirectly assessing biomass accumulation and hydraulic deficiencies. Long term studies using these methods resulted in a new anaerobic reactor design which has been utilized at full scale at Hardee Farms, Lambeth, Ontario, and Lakeview WPCP, Mississauga, Ontario.

Introduction

The development of novel high solids retention time designs for biological reactors has resulted in many new anaerobic systems which are suitable for the treatment of concentrated organic wastewaters. In contrast to conventional treatment alternatives these anaerobic processes can operate with minimal or negative energy requirements, and greatly reduce the quantities of sludge generated for ultimate disposal.

The recent evolution of high rate anaerobic technology has resulted from the recognition that advanced biological reactors require optimization in at least four process design areas. These include:

- microbial kinetics;
- reactor biomass concentrations;
- reactor mass transfer characteristics; and,
- hydraulic retention time (HRT).

Biological kinetics are often beyond the realm of the design engineer as they are controlled primarily by the nature of the wastewater and the make-up of the microbial population which develops spontaneously in the process. These factors can be ameliorated somewhat through manipulation of the reactor environment. In the case of anaerobic treatment, several researchers have indicated that process efficiency can be altered by changes in temperature, pH, oxidation-reduction potential, and the addition of selected micro- and macro-nutrients.

Most of the new anaerobic reactor designs attempt to maximize reactor biomass concentrations and mass transfer to permit a reduction in the required HRT. Many potential users of anaerobic technology believe that these objectives are at odds. As higher biomass concentrations are achieved in a reactor of specified volume, the mass transfer characteristics may deteriorate and the actual HRT may be reduced to impractically low values. This concern has been borne out by literature reports of problems with plugging and channelling in laboratory scale anaerobic filters (Friedman et al. 1981; Young and Dahab 1983) and in full scale anaerobic packed bed systems (Taylor and Burm 1973; Sheridan 1982).

At Environment Canada's Wastewater Technology Centre, several commercially available high rate anaerobic pilot plants have been operated under controlled experimental conditions for over 4 years (Hall and Melcer 1984). One objective of this program was to compare the impact of long term biomass accumulation on the performance of the individual reactors. Selected results from two of the systems studied are presented here to illustrate potential weaknesses in their designs. These potential difficulties have been avoided in two recently-constructed full scale anaerobic facilities which utilize a hybrid reactor design.

Materials and Methods

Pilot Plant Facilities

The two processes assessed during this study were an anaerobic filter (AF) and an upflow anaerobic sludge blanket (UASB) reactor. Table 1 contains information on the design characteristics of both pilot plants.

Table 1. Anaerobic Pilot Plant Characteristics

Reactor	Support Medium	Volume (litres) Initial Void	 Empty Bed
AF	9 cm Norton Actifil	900	950
UASB	None	1100	1100

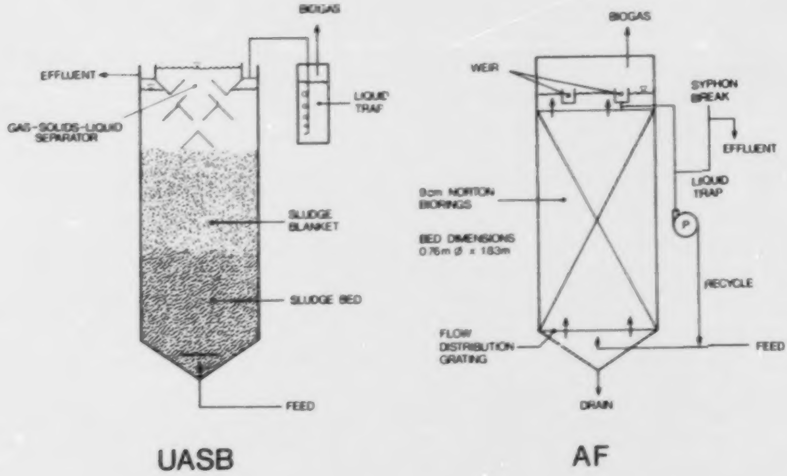

Fig. 1 Pilot scale upflow anaerobic sludge blanket (UASB) reactor and anaerobic filter (AF).

In the AF reactor (Figure 1) wastewater is passed upward through a submerged bed of random plastic packing media. Biological solids produced by growth become attached to the media surface or trapped by settling within the interstices of the bed. The specific surface area provided by media of this type is relatively low (114 m^2/m^3), but since the process retains both attached and non-attached growth, the very high biomass concentrations produced can at least partially offset the mass transfer limitations of this configuration. The packing medium has a high initial void fraction (0.95) which maximizes the available treatment volume as long as excess biomass accumulation is avoided.

The UASB process operates entirely with non-attached biomass and therefore requires no packing medium (Figure 1). As originally described by Lettinga et al. (1979), the successful performance of the UASB depends on the development of "granules" of anaerobic micro-organisms. These granules of biomass can develop to diameters of 1-2 mm, producing particles with very high settling velocities. This allows the biomass to be retained by sedimentation in a sludge bed at concentrations of 40-100 g VSS/L. Above the sludge bed is a blanket zone of more diffuse growth with lower particle settling velocities. Biological COD removal occurs throughout the bed and blanket zones. The biogas bubbles which are produced during anaerobic treatment in the bed are removed by a submerged gas separator. In the region above the gas separator, solid-liquid separation takes place in a quiescent settling zone. All surfaces in the settling zone are constructed with steep slopes to allow settled solids to return to the bed-blanket regions.

The operating conditions maintained in the two pilot plants varied widely over the course of the 4 year study depending on the characteristics of the wastewaters being treated and the experimental variables under study. Table 2 provides a summary of the range of operating conditions used. The initial reactor start-up procedure has been described previously (Hall and Jovanovic 1983).

Table 2. Anaerobic Reactor Operating Conditions

Parameter	Typical Range
Influent COD (mg/L)	5 000 - 36 000
HRT (days)	0.5 - 2.0
Organic Loading (kg COD/m^3·d)	5.0 - 30.0
Biogas Production (m^3/m^3·d)	0.5 - 6.0
Recycle/Feed (m^3/m^3)	0 - 10.0

Assessment of Reactor Hydraulic Efficiency

It was assumed that over a long term period of operation, the accumulation of biomass might lead to changes in the internal hydraulics of both the AF and UASB reactors. Specifically, biomass-associated dead volume was expected to increase. Flow channelling or bypass was also expected to occur if biomass concentrations became excessive.

Quantitative estimates of these hydraulic characteristics were made periodically by utilizing a tracer response technique coupled to non-linear modelling of the flow. Details on the methodology of the residence time distribution (RTD) analysis technique are provided by Hall (1984).

Reactor Biological Solids Measurement

The biological solids mass in the UASB was estimated frequently by drawing samples from equally spaced sampling points covering the entire depth of the reactor. Volatile suspended solids (VSS) concentrations in each sample were then used to calculate the total biomass in the UASB. At less frequent intervals, the UASB was drained into an open tank and samples were collected for direct measurement of the total biomass.

In the AF, non-attached biomass was measured by draining the reactor contents into an open tank for solids sampling. Usually, lightly-attached solids were included in the measurement by repeating the procedure after re-filling the AF with reactor effluent which had been stored for this purpose. The rinsing liquid was also collected in an open tank and additional samples taken for VSS analysis. All non-attached solids were usually discarded after measurement.

After draining and rinsing the bed, the fraction of the initial reactor void volume which could not be recovered by draining was assumed to be a stagnant or dead zone occupied primarily by attached biofilm. Samples of the biofilm were scraped from the medium at several locations in the bed and their volumes were measured by displacement of distilled water. The volumetric solids concentration of these samples were determined as total volatile solids. The average solids concentration was then multiplied by the assumed reactor biofilm volume to obtain an estimate of the total attached biomass.

Pseudo-Steady State Experiments

At several points in the overall 4 year program, a set of pseudo-steady state experiments was carried out to study the short term response of the reactors under different modes of operation. One such study was undertaken in the third year to examine the effects of organic loading and effluent recycle. A factorial design approach was used to establish the combinations of loading and recycle to feed ratio (R/F) to be utilized in a randomized series of two-week runs. The experimental design was a simple 2^2 factorial with centerpoint replicates for variance estimation.

Table 3. Pseudo-Steady State Experimental Design

Run	Design Settings	
	Organic Loading (kg COD/m^3·d)	R/F (m^3/m^3)
1	12.5	4
2	20.0	8
3	5.0	0
4	12.5	4
5	5.0	8
6	20.0	0

At the beginning of each run, the reactor operation was adjusted to the settings specified by the experimental design (Table 3) and process performance was assessed by daily monitoring of the effluent and biogas. Organic loading was controlled by manipulation of the influent flow rate. With the exception of the two study variables, the general operating conditions were the same as those described previously by Hall and Jovanovic (1983). The wastewater utilized during this pseudo-steady state experiment was thermal sludge conditioning liquor obtained from the Lakeview Water Pollution Control Plant in Mississauga, Ontario (Table 4).

Sampling and Analysis

Where possible, samples taken for chemical analysis were 16 hour composites. Sample preparation and analysis were generally performed as described in Standard Methods (1980). Chemical oxygen demand was determined by the culture tube method of Knechtel (1978).

Table 4. Thermal Conditioning Liquor Characteristics

Parameter	Level (mg/L) 50%*	Level (mg/L) 90%*
COD	11 500	15 100
COD-filtered	10 300	13 800
TKN	870	1 120
Total P	56	78
NH_4-N	320	450
TSS	300	880
VSS	270	660
Volatile Acids		
C_2	720	1 240
C_3	370	720
C_4	120	250
pH	4.5-5.0	

*Percent of observations less than value indicated.

Results and Discussion

Volatile Suspended Solids Loss in AF and UASB

During the execution of the pseudo-steady state experiments described in Table 3, anaerobic reactor influent flow rates and effluent recycle rates were varied over wide ranges in a randomized fashion (Figure 2a & b). Prior to the start of the first run on day 87, the AF and UASB systems had been operating for approximately 60 days at an organic loading of 10 kg COD/m^3·d and an R/F of zero. On day 87, the loadings were increased to 12.5 kg COD/m^3·d and effluent recycle was initiated at an R/F of 4.0. Subsequent runs followed the experimental design of Table 3.

The performance of the two anaerobic reactors is partially illustrated by Figure 2c in which COD removal efficiencies are plotted for the pseudo-steady state experimental period. The thermal conditioning liquor was essentially a soluble wastewater in which approximately 80% of COD was biodegradable. Percent COD removal was calculated using unfiltered influent and filtered effluent samples. Although the UASB exhibited slightly lower removal efficiencies than the AF for most of the experimental period, their performances were comparable until day 160. COD reduction averaged 70% in the AF and 60% in the UASB prior to day 160.

Run 6 commenced on day 159. In the AF reactor, COD removal temporarily decreased to 40% before recovering to the 55% level. The UASB system was not operated under the high load conditions specified for run 6. Instead it was maintained in the previous low load mode. On day 164 the UASB was taken out of service to avoid complete process failure due to excessive biomass washout.

At the beginning of the pseudo-steady state runs the total biomass quantities in each reactor were estimated at 20 kg VSS. The estimates were obtained by direct solids sampling in the UASB, and RTD/draining measurements in the AF. For the UASB, essentially 100% of the biomass could be considered non-attached. In the AF, 40% of the total VSS content was non-attached. An examination of the non-attached solids in both reactors indicated that their physical properties were similar. With thermal conditioning liquor as the influent, no sludge granules were detected. However, precipitation of iron and calcium salts produced a large inorganic fraction which reduced the VSS/TSS ratio to less than 0.50, and resulted in a diffuse gritty sludge.

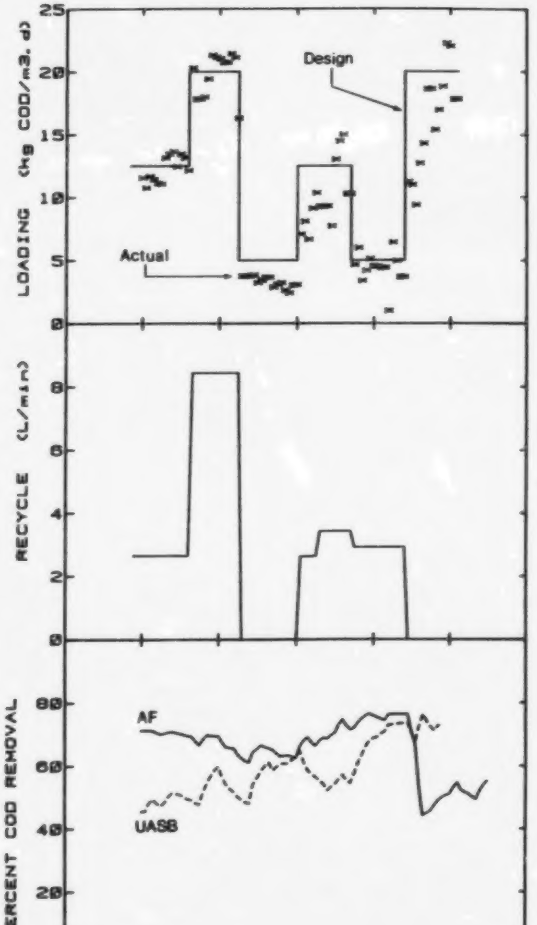

Fig. 2 Operating conditions and observed COD removal efficiencies during pseudo-steady state experiment.

Although both reactors contained similar quantities of non-attached biomass prior to the pseudo-steady state experiment, their responses to the imposed changes in organic loading and R/F ratio were very different. Figure 3 shows the VSS concentrations measured in the AF and UASB effluents during the experiment. On day 88, immediately following the start-up of the recycle pump, both reactors began to lose solids in their effluents. For the AF, effluent VSS levels increased to 1500 mg/L for the duration of the first run, and then declined to an average of 500 mg/L. This level was maintained for the remainder of the runs even though several additional operating changes were imposed on the AF. The UASB exhibited a far more critical solids loss under the same conditions. Effluent VSS rapidly increased to 6000 mg/L during the first run, and never fully recovered. By day 164, the quantity of biomass remaining in the UASB was so minimal that the reactor was taken out of service.

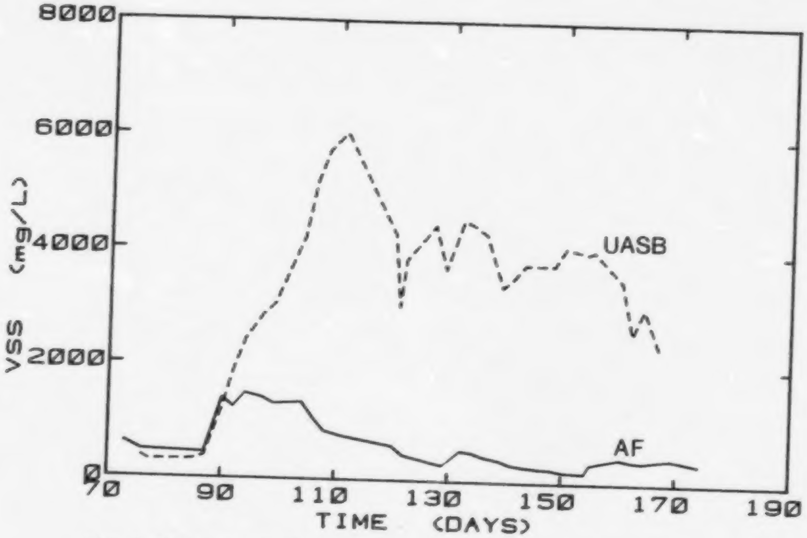

Fig. 3 Effluent volatile suspended solids during pseudo-steady state experiment

The contrasting responses of the two systems indicated that for the case where a non-granular anaerobic sludge is produced, a high rate reactor with a bed of random packing provides superior process stability. The presence of a biofilm which is physically attached to the packing surface prevents complete washout of the biomass. More importantly, the packed bed itself also appears to reduce the loss of non-attached solids during transients.

The mechanism of non-attached biomass retention in a packed bed anaerobic reactor is not well understood. Young and McCarty (1968) observed that rising gas bubbles in an AF reactor tended to carry sludge particles upward. However, after contacting an overlying fixed surface, the bubbles and particles separated. The particles then settled to their original positions as the gas bubbles continued to rise. Assuming that media contact is the controlling mechanism, the use of a random packing would maximize the contact opportunities for rising gas and sludge particles. On the contrary, modular media which are more vertically-oriented may be less efficient in retaining non-attached anaerobic biomass.

Hydraulic Studies

Tracer response techniques were used to assess the long term changes in the mixing characteristics of the two anaerobic reactors. By fitting a hydraulic model to the tracer results, estimates could be made of Qa, the active flow, and Va, the reactor active volume. The fraction of total influent flow defined as active, is that portion which does not bypass the working treatment zone of the reactor. The active volume is the portion of initial reactor voidage which remains actively mixed and therefore constitutes the working zone of the system. In a high rate anaerobic reactor it is likely that both Qa and Va will decrease if biomass accumulation produces plugging and flow channelling.

The results of several tracer studies carried out over the entire four-year program are shown in Figures 4 and 5 for the AF and UASB reactor respectively. The estimated biomass quantities in each reactor are also indicated. In Figure 4, the biomass has been partitioned into its two components, attached VSS (AVSS) and total VSS (TVSS). The sawtooth pattern results from wasting of non-attached VSS initially, and then also a portion of the AVSS on days 1068 and 1213. In Figure 5, all biomass shown in the UASB is as non-attached VSS. Solids wasting was never required in the UASB.

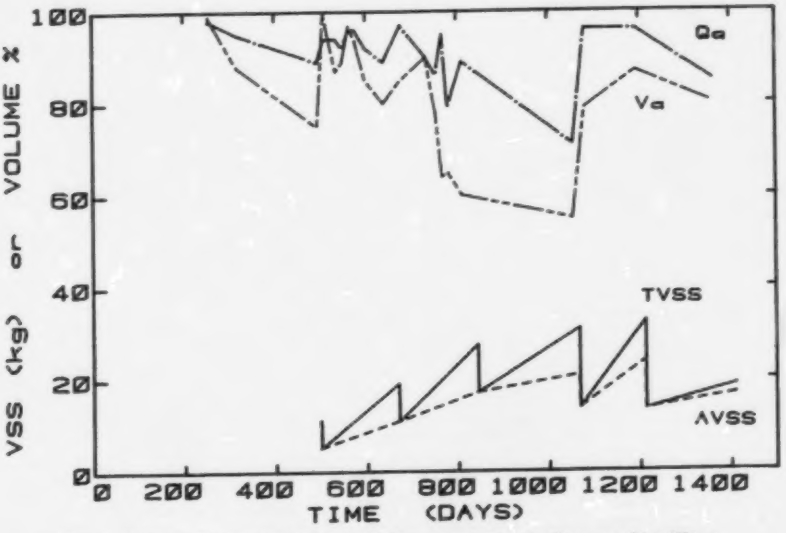

Fig. 4 Mixing characteristics and biomass content of anaerobic filter

In the AF, Qa and Va decreased significantly during the first 3 years of operation (Figure 4). Periodic removal of the unattached VSS helped to reduce the rate of this decrease, but it was insufficient to completely reverse the effect. Consequently, by day 1054, the active volume was reduced to 55% of the initial voidage and approximately 30% of the incoming flow was bypassing the bed. At this point several unsuccessful attempts were made to remove a portion of the attached biofilm using high velocity effluent recirculation and/or biogas sparging below the bed. On day 1069 it was found that draining of the bed, followed by effluent irrigation from the top, was successful in removing about 35% of the biofilm attached to the media. This produced an immediate improvement in AF hydraulics with no significant impact on the treatment performance of the reactor. The same upgrading approach has been used subsequently with satisfactory results.

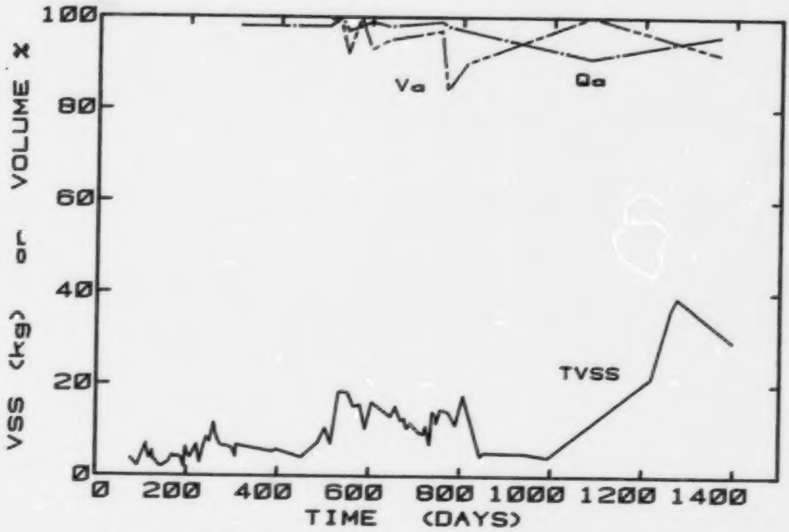

Fig. 5 Mixing characteristics and biomass content of UASB

During the same four-year operational period, very similar quantities of biomass were measured in the UASB (Figure 5). However, tracer studies indicated that the UASB active volume never constituted less than 80% of the initial void volume. In fact, the active volume percentage was usually greater than 90%. Similarly, bypass flow averaged only 3% of the influent rate.

A comparison of the data in Figures 4 and 5 indicated that with comparable biomass inventories, the UASB configuration provided better control over internal reactor hydraulics. Essentially 100% of the reactor liquid volume remained actively mixed and very little bypass flow was detectable. With a similar quantity of retained biomass, the presence of media in the AF caused a severe deterioration in the reactor mixing characteristics over a long period of operation. These hydraulic problems could be addressed only by reactor shutdown for the removal of excess biomass. This remedial procedure also appeared to be required on a regular basis.

Hybrid Reactor Configuration

From the preceeding discussion it is apparent that the AF and UASB reactor configurations provide both some distinct advantages for anaerobic treatment, as well as potentially serious disadvantages. To overcome the weaknesses associated with the individual process designs, several researchers have proposed combining basic elements of each in a novel hybrid anaerobic reactor system. The result is a unit which is designed as a UASB in the lower part of the reactor, and as an AF in the upper part.

The concept of a hybrid anaerobic reactor was introduced by Olthof and Oleszkiewicz (1982) who discussed its application for industrial wastewater treatment. Lettinga et al. (1983) provided an overview of the advantages and disadvantages of a combined UASB-AF process. More recently, Guiot and van den Berg (1984) reported that a laboratory scale hybrid reactor was able to treat organic substrates at loadings up to 25 kg COD/m^3·d with 90% COD conversion efficiencies. The same authors found that the plastic rings utilized in the packed zone provided an efficient mechanism for biomass retention in the sludge blanket zone.

As a result of the operating difficulties associated with UASB and AF reactor designs, and the positive results obtained at laboratory scale by Guiot and van den Berg, the 1 m^3 AF pilot plant at the WTC was converted to a hybrid configuration (Figure 6). The only change from the previous design was the relative volume of the packed zone. In the WTC hybrid reactor, the packed zone occupies the upper 45% of the empty bed volume - leaving the lower 55% for accumulation of non-attached sludge blanket biomass. The hybrid reactor was started with a pulp and paper wastewater in June 1984.

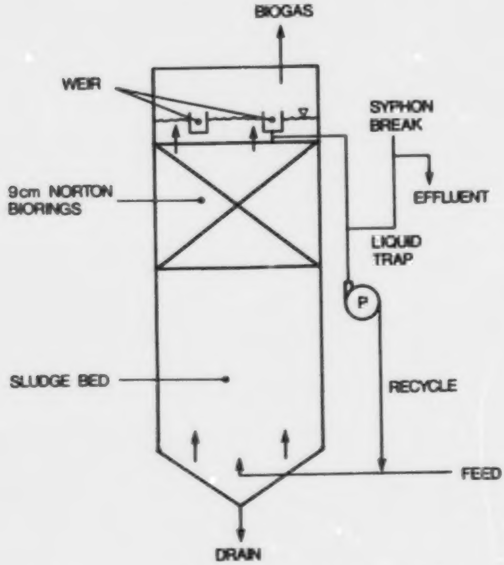

Fig. 6 Schematic of pilot scale hybrid anaerobic reactor

The absence of media in the lower part of the hybrid process provides the potential for eliminating the hydraulic inefficiencies observed with the packed bed AF. As indicated by the UASB tracer studies, the sludge blanket zone should remain well mixed with essentially 100% volume utilization. Without media in this zone, the blanket can be easily monitored and biomass wasting can be accomplished in a straightforward manner. By allowing more than 50% of the reactor volume for non-attached solids accumulation, the sludge bed is free to develop to the 1-2 m heights reported by Lettinga et al. (1983) for high capacity UASB reactors. With 1-2 m of biomass in the lower reactor zone, the hybrid process will undoubtedly accommodate elevated organic loadings irrespective of the contribution from the packed zone.

When the hybrid reactor is applied to wastewaters which are conducive to biomass granulation, the lower zone sludge bed will be easily retained. However with some wastewaters, granulation of the anaerobic sludge proceeds slowly, and the resulting biomass has poor settling characteristics. In this situation, a zone of random packing in the upper part of the hybrid reactor will improve the retention of the non-attached biomass as illustrated in Figure 3.

The packed zone confers several other advantages on the hybrid reactor design. As noted by Hall and Jovanovic (1983), upflow randomly packed beds may exhibit shorter start-up times than UASB systems. This difference is most significant when sludge granulation does not proceed satisfactorily. The packed zone can also develop a fixed biofilm which can contribute to the total treatment capacity. Since this biomass is physically attached to the media it provides an extra element of process stability under transient conditions. In the event that serious process upset occurs, the fixed biomass, can serve as a seed source for reactor recovery or re-start. As this biofilm accumulates, the hydraulic inefficiencies shown in Figure 4 will also increase. However, the overall significance of these inefficiencies will be reduced in proportion to the reactor volume fraction occupied by the packed zone. With a high capacity sludge blanket zone below it, plugging of the packed-zone may have only a minimal impact on the overall reactor performance. Excess biofilm can be removed from the packing by draining the hybrid reactor contents to a point below the bed, followed by irrigation of the packing from above.

Perhaps the most significant advantage of the hybrid configuration is the reduced cost of the support media required. Commercially available packing materials contribute between Cdn. \$100 and \$300/m^3 to the capital cost of a high rate anaerobic reactor. By reducing the volume of media used by 50% - 75%, the process economics become decidedly more attractive.

Full Scale Hybrid Anaerobic Systems

Design information and full scale performance data for hybrid reactors are not readily available at this time. Young and Dahab (1983) have shown that both modular and randomly packed media can effect the retention of attached and non-attached biomass, but no rigorous comparison of the two types was carried out. At the University of Manitoba, studies are underway to assess the effects of varying packed zone depth and volume ratios, as well as the operating strategies most applicable to hybrid systems. Even though detailed process design data are not yet available, two Canadian full scale anaerobic treatment projects are proceeding with hybrid reactor designs. For these two applications, it was decided that the potential advantages of the hybrid process outweigh the risks associated with a new reactor design.

At Hardee Farms International in Lambeth, Ontario, two parallel hybrid reactors have been constructed for anaerobic pre-treatment of vegetable processing wastewater (Love, 1984). Both tanks contain modular plastic media located above a shallow zone for non-attached solids accumulation (Figure 7). Different media depths were used in the two reactors to provide comparative data on the effects of bed depth. The anaerobic facility was seeded with digested manure in November 1983, and was operating at 65% of design hydraulic capacity prior to the seasonal production shutdown in February 1984. The system was restarted in June 1984 and has operated as close to design conditions as permitted by wastewater availability.

A second hybrid facility is under construction at the Lakeview Water Pollution Control Plant in Mississauga, Ontario, for the treatment of thermal sludge conditioning liquor. Anaerobic treatment of this TCL source was studied extensively in laboratory scale AF units (Crawford et al. 1982), and at pilot scale utilizing several different reactor configurations (Hall and Jovanovic 1983). It was found that in packed bed reactors treating TCL, the attached and non-attached solids included large quantities of inorganic material which originated from precipitation of wastewater constituents. These inorganic solids resulted in dense biofilms which could plug the packing in an AF reactor. In addition, pilot scale UASB studies indicated that the non-attached solids did not granulate, and therefore, were difficult to retain in a system without media.

To minimize the consequences of these two problems a hybrid configuration was selected for full scale treatment of 1400 m^3/d of TCL (Crawford 1984). Two existing anaerobic sludge digesters are being refitted with influent distribution and media support systems (Figure 8). A relatively deep sludge bed zone has been incorporated below a 2 m packed bed zone which will utilize random plastic media. Start-up of the two parallel hybrid reactors is scheduled for late 1984.

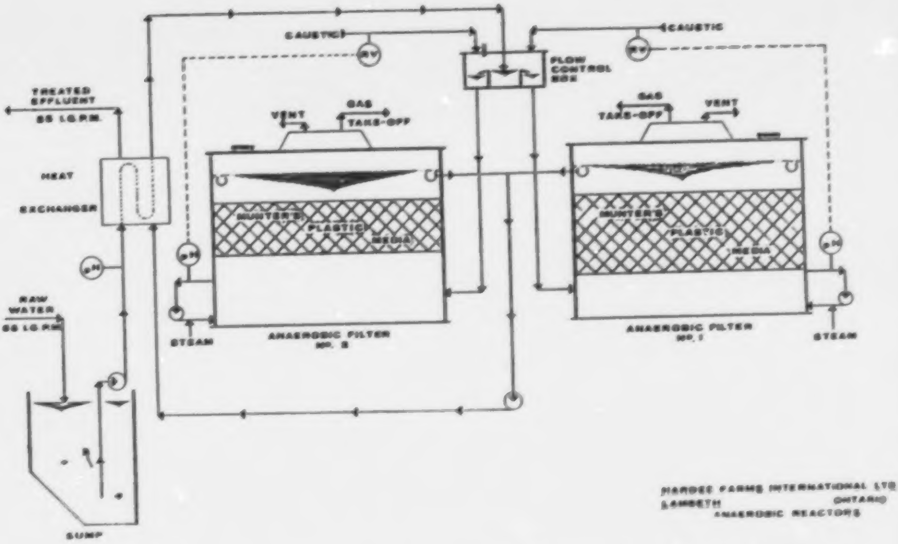

Fig. 7 Anaerobic treatment system at Hardee Farms, Ontario (Love, 1984)

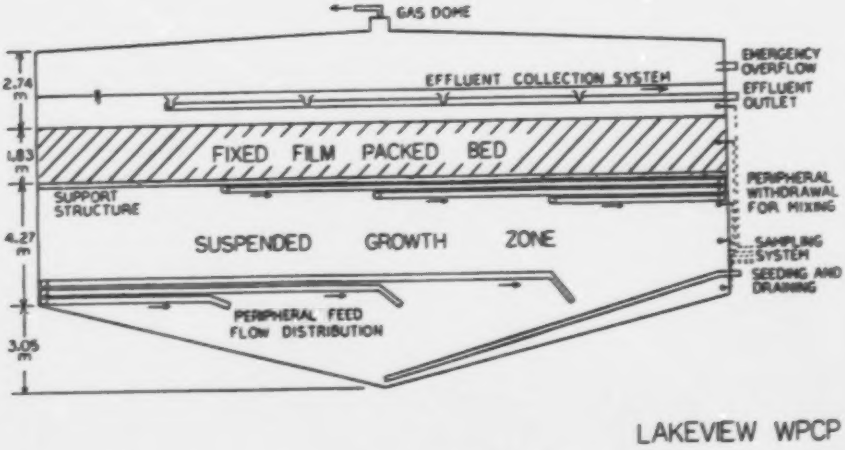

Fig. 8. Schematic of anaerobic reactor design for Lakeview WPCP, Mississauga, Ontario (Crawford, 1984).

Summary and Conclusions

1. During transient operation with a non-granular sludge, the UASB process was susceptible to severe biomass washout which led to process failure.

2. Under similar operating conditions, the presence of a random packing medium in an anaerobic filter prevented the loss of non-attached biomass.

3. The anaerobic filter demonstrated a long term deterioration in internal hydraulic characteristics resulting from excess biomass accumulation. This deterioration could be counteracted by periodic removal of both attached and non-attached solids from the reactor.

4. The UASB process exhibited satisfactory hydraulic characteristics under all conditions examined.

5. A hybrid anaerobic reactor configuration is described which could eliminate the disadvantages associated with the operation of either the UASB or anaerobic filter.

6. The hybrid reactor concept has been utilized in two Canadian full-scale anaerobic treatment applications.

Acknowledgements

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ANAEROBIC TREATMENT OF CHEESE WHEY

by

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Introduction

In 1979, 41% of the cheese produced in Canada was manufactured in Ontario (1,2). For each kilogram of cheese manufactured, approximately 9 kilograms of whey is also produced. Cheese whey is a warm, high strength organic waste product which can create disposal problems in provinces such as Ontario where there are a large number of cheese producers. The most satisfactory method of whey disposal is to re-use the organic components in one of several options available for resource recovery (1,2). Unfortunately, the supply of whey for re-use often exceeds the demand. This situation has left many smaller cheese plants in an insecure position with regard to whey disposal, since economically acceptable alternatives to the re-use options have not been identified.

Anaerobic methane fermentation offers several potential advantages for the disposal of whey. The major benefit to small cheese plants is that most of the organic matter in the whey could be recovered as a fuel grade biogas. In addition, by treating the whey at the plant site, the costs of transportation to a centralized disposal facility would be eliminated. The final product of the process would be a much lower strength wastewater which might be suitable for further treatment or disposal using more traditional means.

Several studies have shown that anaerobic treatment can achieve adequate removals of chemical and biochemical oxygen demand (COD and BOD) from cheese whey, while producing methane at close to theoretical yield values (3-12). Most of these reports proposed the use of fixed film anaerobic reactors to obtain the benefits of high solids retention time (SRT) and short hydraulic retention time (HRT) operation (3-8). However, since many of the published studies were carried out using highly diluted whey (3,5-7, 10), or in very small laboratory scale reactors (3,5,7-10), the results have limited significance for the design of a full scale treatment system.

To provide design data for anaerobic treatment of cheese whey, Agriculture Canada and Environment Canada sponsored a pilot scale study at the Wastewater Technology Centre (WTC) using a downflow stationary fixed film (DSFF) reactor. This paper summarizes the results of the study and discusses the implications for process selection and performance of a full scale facility which is currently operating in Ontario.

Materials and Methods

The pilot scale anaerobic facility used in this study was a 520 liter downflow stationary fixed film (DSFF) reactor designed on the principles described by van den Berg and Kennedy (8). The reactor incorporated a 2.0 m deep bed of Cloisonyle plastic media which produced a large number of vertical tubular channels about 1.5 cm in diameter, with an initial voidage of greater than 95%. Influent to the DSFF was pumped through an in-line electric heater which adjusted the wastewater temperature to 35°C. The reactor was operated in the downflow mode with the bed fully submerged. A portion of the effluent was pumped through a flow cell for online pH measurement, followed by recirculation to the reactor inlet. The reactor pH was controlled at 6.5 by the automatic addition of 1 to 2 N NaOH to the influent.

Several batches of whole cheese whey were delivered to the WTC by refrigerated tank truck and stored in a cooled, 70 m^3 tank. The average concentrations of several constituents of the whole whey are shown in Table 1. Feed for the DSFF was prepared by combining whey and tap water in a refrigerated 5000 L tank. Nutrients were not added to the final reactor feed. Since whey was often stored for periods of 2 - 3 months, bacterial growth was unavoidable. The average suspended solids level observed for the whole whey probably reflects the limited growth that occurred during storage.

Table 1. Characteristics of Whole Whey Used in Pilot Study.

Parameter	Average Concentration (mg/L)
COD	61 000
COD-filtered	57 000
BOD	26 000
BOD-filtered	25 000
TKN	980
NH_4-N	155
Total-P	510
Suspended Solids	1 780
Volatile Susp. Solids	1 560

The DSFF reactor was seeded with municipal digester sludge and started with a combination of whey and thermal sludge conditioning liquor as the substrate. After 240 days of operation, treatment of 100% whey had been achieved at an organic loading of approximately 12.5 kg COD/m^3·d. For the remaining 12 months of the study, the reactor was operated under pseudo-steady state conditions at several organic loading rates between 4.0 and 17.5 kg COD/m^3·d with influent whey concentrations of 100% and 50%. The 50% strength feed represented the wastewater concentration expected at many Ontario cheese plants, where washwater could be combined with the raw whey prior to anaerobic treatment. Each pseudo-steady state experimental run was at least one month in duration.

Samples of reactor influent and effluent were collected three to five times per week for chemical monitoring using the sample preparation and analysis techniques proposed in Standard Methods (13). Biogas flow rates were measured in oil-filled wet test meters. Gas chromatographic procedures were used for volatile acids and gas quality determinations.

Results

The pseudo-steady state treatability runs completed with 100% cheese whey indicated that very high conversion efficiencies could be obtained with anaerobic treatment. Figure 1 shows that at loading rates between 6 and 18 kg COD/m^3·d, soluble and total COD removal averaged 95% and 75% respectively. Since the COD in cheese whey is almost entirely soluble (Table 1), the removal efficiency of soluble COD represents the biological COD conversion achieved in the DSFF process. Total COD removals are lower due to the presence of biological solids which leave the reactor with the treated effluent.

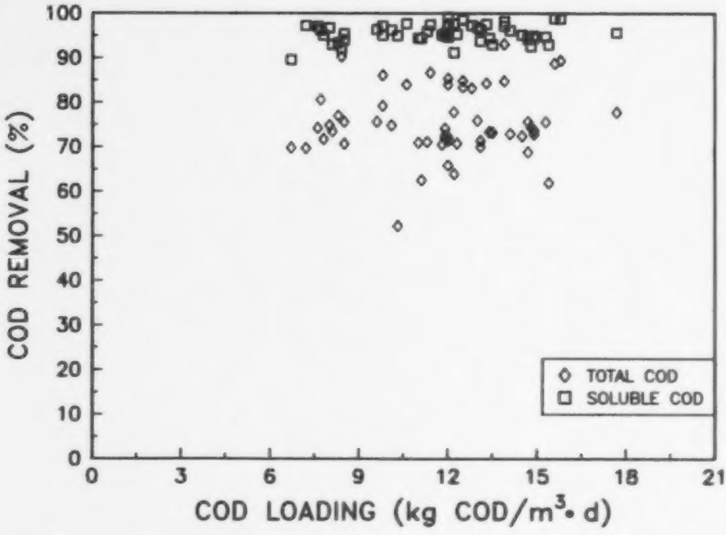

Fig. 1. COD removal efficiencies observed during anaerobic treatment of 100% whey in pilot scale DSFF reactor.

Table 2 summarizes the performance of the DSFF and the average effluent quality produced during operation at two pseudo-steady state loading rates with 100% whey. The observed BOD removal efficiencies were similar at both the high and low loading rates and compared favourably to the COD removals shown in Figure 1. Relatively high effluent COD concentrations were noted even though satisfactory overall COD removal efficiencies were obtained. Most of the effluent COD was due to the presence of about 8600 mg/L of biological solids and 1350 mg/L of volatile acids. Caustic addition required for pH control was in the range of 1.6 to 2.2 g/L of wastewater. A small proportion of the influent organic nitrogen present in whey protein (Table 1) was hydrolysed during anaerobic treatment. This produced effluent ammonium nitrogen levels of 100 to 160 mg/L. Finally, the methane yield measured at the high loading rate was significantly lower (0.28 m^3/kg COD_r) than that observed during the low rate run (0.34). However, both values compared reasonably well to the theoretical methane yield of 0.35 m^3/kg CODr.

Five pseudo-steady state runs were carried out with a feed consisting of 50% cheese whey in tap water. As shown in Figure 2, COD removal efficiencies were lower and more variable than those observed with 100% whey. At the lower loading rates, the soluble and total COD removals averaged 90% and 75% respectively. At loading rates above 12 kg COD/m^3•d the corresponding values were as low as 80% and 64%. The BOD removal efficiencies shown in Table 3 varied in a similar fashion to the COD removals. Soluble and total BOD removal ranged from 76% to 90%, and from 64% to 87%. The measured effluent VSS levels were approximately one-half of those observed with 100% whey. These lower values are the result of operation with a more dilute influent. However, because the volatile acid levels were much higher (1400 to 3560 mg/L) than those observed with whole whey, the effluent total COD concentrations were higher than expected for the dilute feed. Caustic requirements were also correspondingly higher (0.8

Table 2. Pilot Plant Performance and Effluent Quality with Full Strength Cheese Whey Feed

Parameter	Pseudo-Steady State Value	
COD Loading (kg/m^3·d)	13.0	8.3
Retention Time (d)	4.9	6.6
Total BOD Removal (%)	82	84
Soluble BOD Removal (%)	93	92
Effluent COD (mg/L)	15 300	13 500
Effluent TVA (mg/L)	1 350	1 380
Effluent VSS (mg/L)	8 620	8 610
Effluent NH_4-N (mg/L)	164	100
CH_4 Yield (m^3/kg COD_r)	0.28	0.34

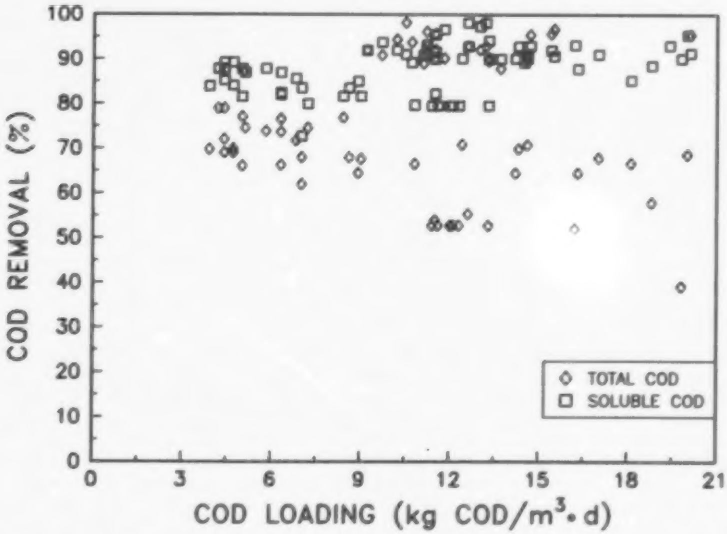

Fig. 2. COD removal efficiencies observed during anaerobic treatment of 50% whey in pilot scale DSFF reactor.

to 3.0 g/L wastewater). With the exception of the first and last runs at relatively high loading rates, the observed methane yield coefficients were again very close the theoretical value of 0.35 m^3/kg COD_r.

Discussion

The treatability results summarized above indicate that at both 100% and 50% strength, cheese whey is amenable to anaerobic treatment in a DSFF reactor. Nonetheless, two major difficulties were identified during the pilot scale study. The first problem resulted

Table 3. Pilot Plant Performance and Effluent Quality with 50% Cheese Whey Feed

Parameter	Pseudo-Steady State Value				
COD Loading (kg/m^3·d)	13.1	15.4	7.4	4.6	12.1
Retention Time (d)	2.7	1.7	4.1	6.1	2.5
Total BOD Removal (%)	82	64	81	87	73
Soluble BOD Removal (%)	89	77	84	90	76
Effluent COD (mg/L)	9480	10805	9620	7400	13700
Effluent TVA (mg/L)	1470	3180	2110	1400	3560
Effluent VSS (mg/L)	5160	4730	2140	1500	5230
Effluent NH_4-N (mg/L)	80	130	130	251	65
CH_4 Yield (m^3/kg COD_r)	0.25	0.35	0.33	0.33	0.26

from the very high volatile acid levels that were observed in the DSFF effluent. Although organic loading rates as low as 4.6 kg COD/m^3·d were applied, it was not possible to reduce the effluent volatile acids to less than 1400 mg/L. At higher loading rates, volatile acids levels of 3000 to 5000 mg/L were measured routinely.

Effluent volatile acid concentrations of this magnitude represented large quantities of readily biodegradable BOD discharged in the treated effluent. Of additional concern was the fact that control of the DSFF pH with these elevated volatile acids levels required the addition of large quantities of alkalinity. A limited amount of alkalinity in the form of NH_4CO_3 was produced by hydrolysis of proteinaceous material in the whey during anaerobic treatment. However, these levels of NH_4 (Tables 2 & 3) were insufficient to maintain reactor pH. Figure 3 illustrates the relationship between additional alkalinity required for pH control and the reactor volatile acid concentration. With even the minimum average volatile acid level of 1400 mg/L identified in this study, approximately 1500 mg/L of caustic was required for pH control. At this level of addition, the chemical cost for treatment of whey was about \$1.20/m^3 of feed.

The overall performance of the DSFF reactor in terms of COD removal rates or biogas production rates was similar to those observed in other studies (3-8). However, the volatile acids concentrations were considerably higher than those previously reported. The unexpectedly high volatile acid levels during anaerobic treatment of cheese whey would be consistent with a limitation in one or more trace nutrients. Kelly and Switzenbaum (7) indicated that the addition of micronutrients reduced the effluent volatile acids levels and enhanced the performance of an anaerobic expanded bed reactor treating dilute whey. Li and Corrado (11) also reported relatively low volatile acids during treatment of a high strength whey with micronutrient supplementation. Unfortunately, this reasoning was contradicted by the observations of van den Berg and Kennedy (8) who demonstrated satisfactory performance of a laboratory scale DSFF reactor without the addition of micronutrients.

The second difficulty encountered in the application of fixed film anaerobic technology for the treatment of a very high strength wastewater is illustrated by Figure 4. With cheese whey, the DSFF effluent suspended solids concentrations were almost directly related to the influent COD concentration. The average VSS yield measured during the study was approximately 0.12 kg VSS/kg COD_r. At the highest influent concentrations applied, Figure 4 indicates that effluent volatile suspended solids levels (VSS) of about 1.0% were obtained. The quality of the effluent discharged from the DSFF was greatly reduced by the presence of these biosolids. Several attempts were made to find an inexpensive method for reducing the VSS levels - without success. The very poor settling and dewatering characteristics of the solids rendered each attempt at post-treatment ineffective. It is worth noting however, that although the anaerobic effluent VSS levels were high, the BOD associated with these solids was much lower than that normally encountered with aerobic biosolids. Figure 5 demonstrates that the COD/VSS ratio of

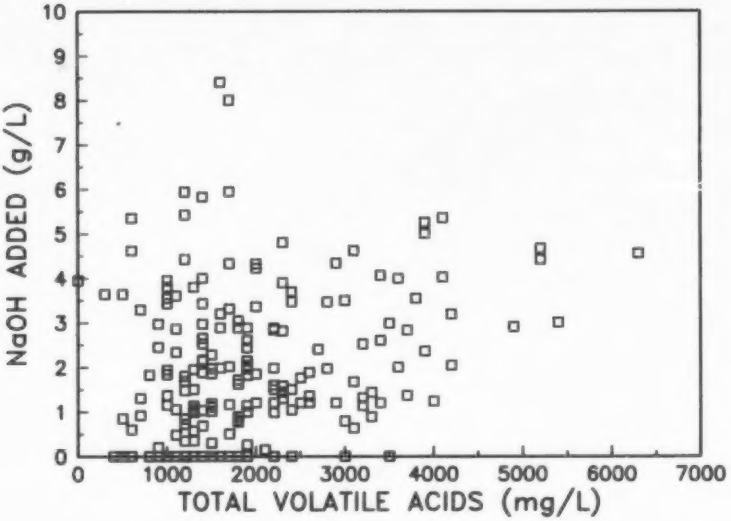

Fig. 3. Relationship between quantity of caustic added for pH control and DSFF effluent volatile acids concentration.

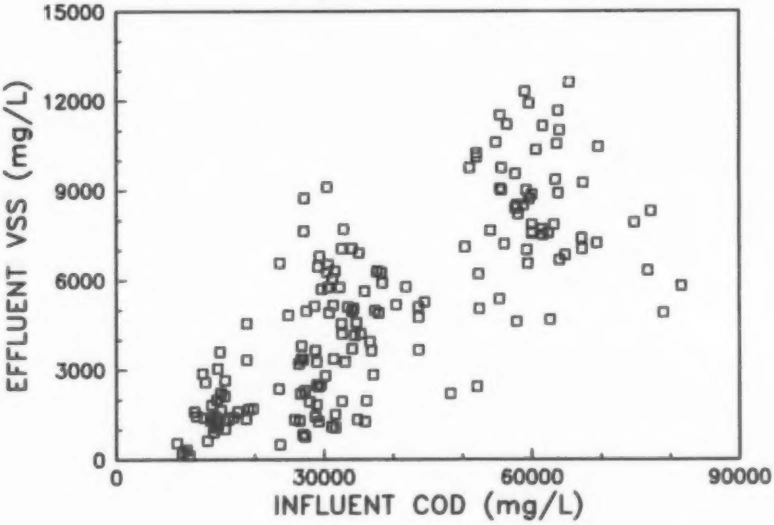

Fig. 4. Relationship between effluent VSS concentration and influent COD concentration observed in pilot scale DSFF.

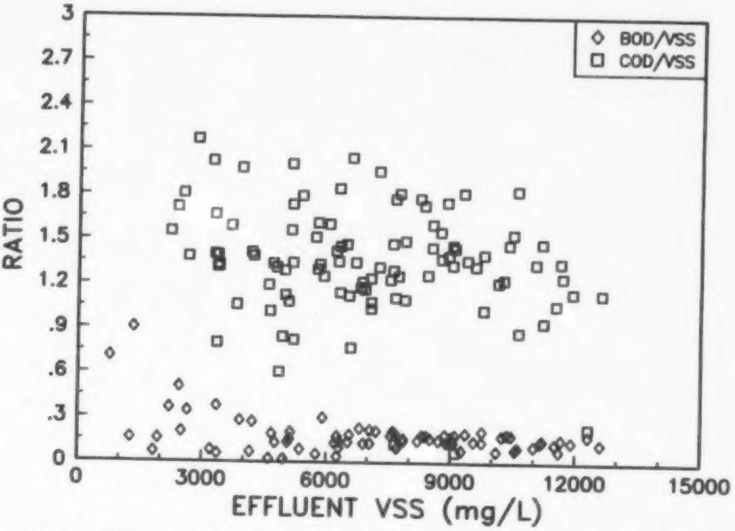

Fig. 5. COD and BOD content of pilot scale DSFF reactor effluent VSS.

anaerobic biosolids is in the usual range of 1:1 to 2:1. The average BOD/VSS ratio of approximately 0.2:1 was about 20% of the corresponding value for aerobic biosolids.

The influence of wastewater concentration on the suitability of anaerobic fixed film technology is often overlooked. The lower practical concentration limit will be set strictly by economic considerations. Certainly it is technically feasible to treat dilute municipal sewage to secondary standards (15). The upper concentration limit for fixed film systems is more difficult to define.

A review of the literature on the kinetics of anaerobic microorganisms (16) indicated that, at 35°C, the doubling time for a mixed anaerobic culture is approximately 2 days. At hydraulic retention times (HRT) above this limit, anaerobic bacteria can reproduce in the liquid phase of a reactor before they are washed out of the system. At HRTs below 2 days, film growth will dominate. Suspended growth can therefore be controlled in a biofilm reactor by operation at short HRTs. For high strength wastewaters such as whey, the maintenance of HRTs of less than 2 days would require unrealistic organic loading rates of greater than 30 kg COD/m^3·d. When suspended growth is favoured, the usefulness of the reactor packing is questionable since the high effluent suspended solids levels will result in a process with characteristics similar to those of a completely mixed suspended growth reactor.

Table 4 summarizes theoretical full scale process designs and costs for a fixed film and a completely mixed reactor for the anaerobic treatment of 50 m^3/d of 100% cheese whey. In the design of the completely mixed reactor, a relatively conservative retention time of 30 days was used. Even so, the comparison establishes that for a high strength wastewater, the fixed film option has no significant cost advantage over a completely mixed system. From a performance perspective, the fixed film design might promote better process stablility - but perhaps also greater concern for flow distribution and media plugging. As shown above, the effluent quality expected from both systems would be similar.

Table 4. Fixed Film and Completely Mixed Reactor Design and Cost Comparison ($ 1983).

Item	Fixed Film	Complete Mix
Whey Volume (m^3/d)	50	50
COD Concentration (mg/L)	65 000	65 000
Retention Time (d)	6.5	30
COD Loading ($kg/m^3 \cdot d$)	10	2.2
Reactor Tank ($)	34 000	115 000
Packing Media ($)	64 000	--
Media Delivery ($)	6 000	--
Media Support ($)	4 000	--
Recycle & Distr'n ($)	14 000	--
Media Installation ($)	4 000	--
Digester Mixing ($)	--	12 000
Additional Eng'g ($)	5 000	--
Insulation ($)	12 000	23 000
Total Cost	$143 000	$150 000

Full Scale Case History

In 1984, Millbank Cheese and Butter Ltd. constructed an 1100 m^3 completely mixed reactor for the anaerobic treatment of whey from a cheese plant at Millbank, Ontario (14). In August 1984, the digester was seeded with 550 m^3 of anaerobically digested beef cattle manure and whey addition was started on an intermittent basis. The performance of the Millbank anaerobic system over the subsequent 2 year period is shown in Figures 6 and 7.

During the first 4 months of operation, the organic loading rate to the digester varied about an average of 1 kg $COD/m^3 \cdot d$ (Figure 6). At this loading rate, the digester retention time was greater than 60 days. Initially, it was anticipated that the reactor would operate without external pH control. However, after the first month, the reactor volatile acids concentration increased steadily to the 5000 - 6000 mg/L range. A reduction in the loading rate to 0.6 - 0.8 kg $COD/m^3 \cdot d$ had only a small effect on the volatile acid levels. Therefore, in January 1985, the decision was made to begin the addition of 50% NaOH to maintain a pH level of 6.5 or greater within the digester. Because of the high volatile acids levels encountered, large quantities of caustic were required to control the digester pH. On several occasions, as much as 8 g NaOH were added per liter of whey treated. This probably resulted in concentrations of Na that were toxic to methanogenic bacteria. As a consequence, effluent volatile acids levels continued to increase until the initial monitoring program was completed in March 1985.

In September 1985, analysis of the digester effluent was resumed. At the same time, caustic addition was eliminated and periodic feeding of digested manure was started. In January 1986 (Day 520), the digested manure was replaced by municipal sludge digester supernatant. The volume of digested sludge added comprised approximately 10% of the total liquid feed to the whey digester. As shown in Figure 6, the digester volatile acid levels decreased slowly over the subsequent 10 month period from 4000 - 5000 mg/L, to 200 - 300 mg/L, while the organic loading rate was maintained at 1.0 kg $COD/m^3 \cdot d$. The steady improvement in digester performance suggests that the supernatant provided either a continuous source of anaerobic microbial seed, or, one or more nutrients that were not present in the whey itself. In the DSFF pilot study, a micronutrient limitation was identified as the most likely cause of the high volatile acid concentration observed.

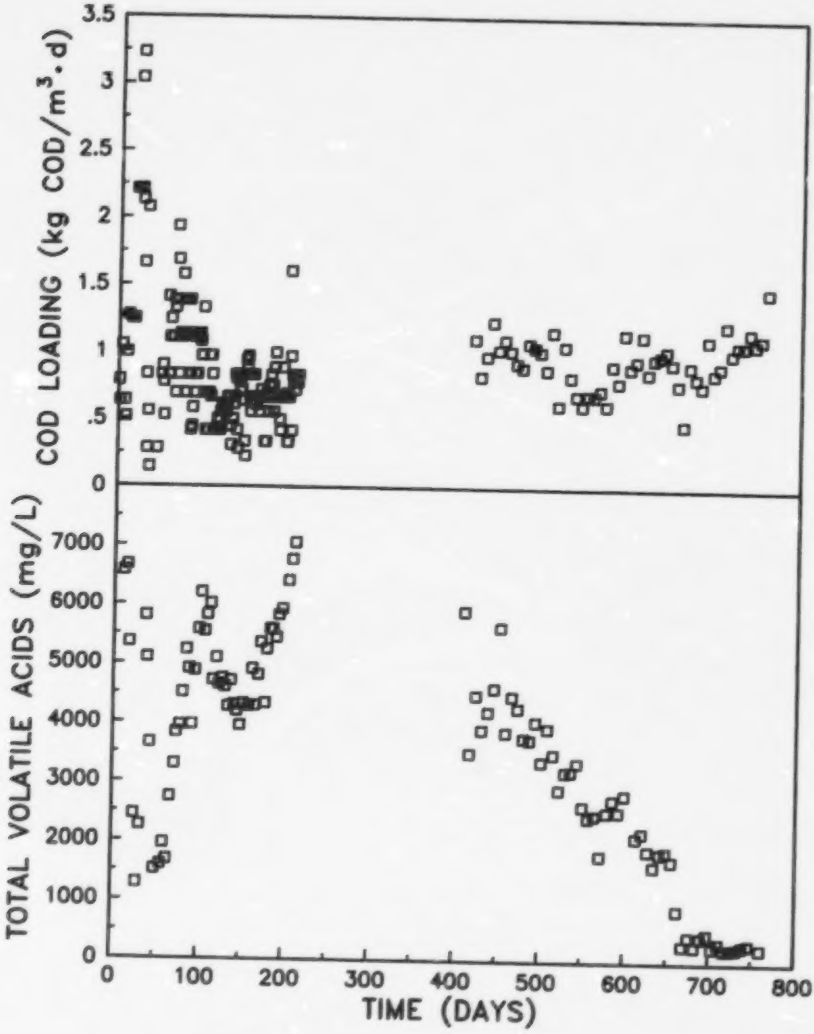

Fig. 6. COD loading rates and effluent volatile acids in full scale completely mixed digester treating 100% whey.

Throughout the entire period of operation, the effluent quality from the full scale digester compared favourably to that achieved with the pilot scale DSFF reactor. Figure 7 indicates that full scale total COD removals were usually between 70% and 90% - the same range as that in the pilot studies (Figure 1). In addition, the effluent VSS levels in the full scale system (4000 to 8000 mg/L) were similar to those observed with 100% whey during the DSFF study. At the end of the monitoring program in August 1986, the digester was operating satisfactorily at approximately 1.2 kg COD/m^3·d. At

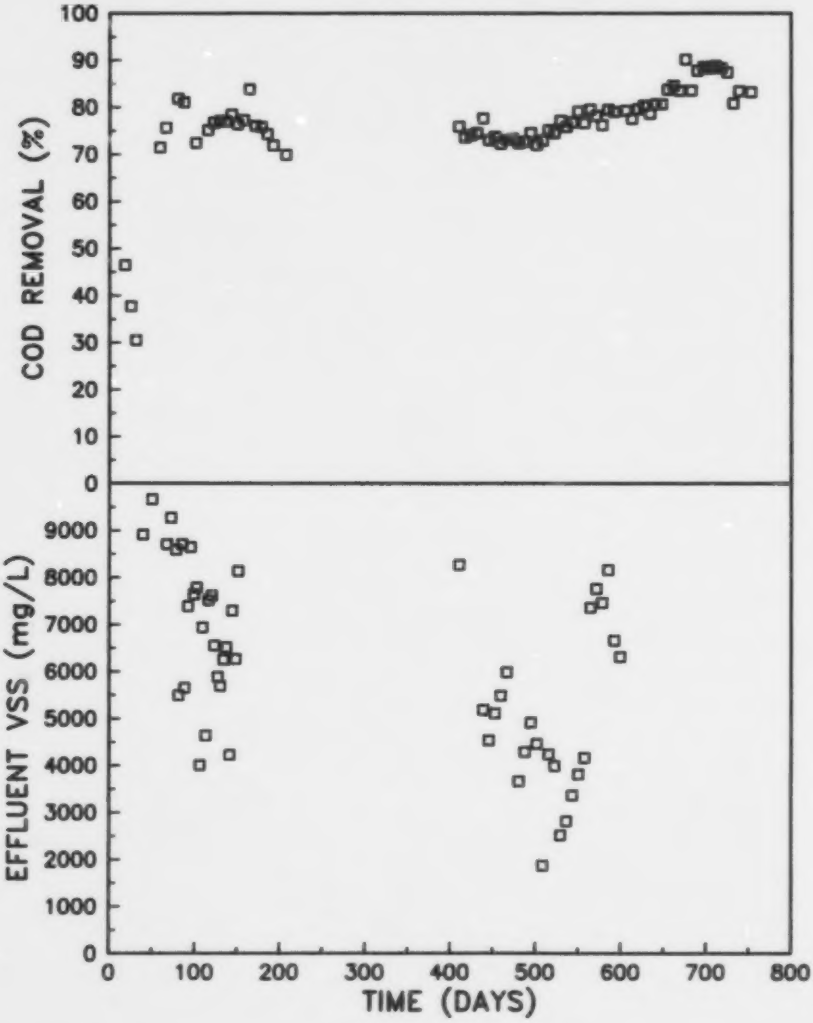

Fig. 7. COD removal efficiencies and effluent VSS concentrations in completely mixed digester treating 100% whey.

this loading rate, the system is treating all of the whey that is currently available for anaerobic digestion.

Summary and Conclusions

1. The BOD and COD content of whole cheese whey could be reduced significantly by anaerobic treatment.

2. A nutrient deficiency in whey may be the primary cause of the high effluent volatile acids observed during both the pilot scale and the full scale studies.

3. Although the soluble COD removal efficiencies observed with cheese whey were satisfactory, the effluent from the fixed film and complete mix reactors contained very high concentrations of biosolids. These biosolids exhibited a low BOD content and poor dewaterability.

4. The generation of a predominantly suspended biomass with very high strength wastewaters resulted in comparable costs for either a complete mix reactor or a fixed film process. With both systems the effluent quality produced by anaerobic treatment was similar.

Acknowledgements

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ANAEROBIC SLUDGE DIGESTER TRACER STUDIES TO ASSESS ALTERNATIVE MIXER CONFIGURATIONS

by

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Abstract

The influence of total mixing power and the frequency of raw sludge pumping on the degree of mixing was investigated at a primary high-rate anaerobic digester using tracer techniques. The digester treated co-settled primary and waste activated sludge and was equipped with a total of 11 gas guns and four 5.6 kW (7.5 HP) compressors. Six residence time distribution (RTD) curves were obtained by injecting lithium chloride into the digester and measuring the lithium concentrations of samples collected from the overflow line between the primary and secondary digesters. Flow short-circuiting and active volume fractions were estimated for different configurations by comparing observed RTD curves to predictions from theoretical mixing models.

Introduction

High-rate anaerobic digesters employ mixing systems to prevent the formation of scum layers, eliminate thermal stratification, and distribute nutrient-rich feed throughout the digester[1]. A well mixed digester achieves a greater degree of volatile solids destruction and produces more gas than a poorly mixed digester[2,3]. Digester mixing systems are commonly designed on the basis of unit power input, the total motor power of the mixers per unit of tank volume[4].

Residence time distribution (RTD) studies provide a rigorous approach for evaluating the performance of installed mixing systems. During an RTD study, a suitable tracer is injected, usually as a pulse input, into the digester and the outlet tracer concentrations are monitored with time. Mixing efficiency is determined from the distribution defined by the outlet tracer concentrations and the times of sample collection. Either a statistical summary of the distribution is determined, or the observed distribution is compared to predictions from mixing models[5].

A number of mixing models are used to evaluate RTD curves. One of the simplest models is the continuous stirred tank reactor (CSTR). For a CSTR injected with a pulse of tracer, the outlet tracer concentration (Ci) decays exponentially with time, Figure 1(a). A plot of the natural log of the tracer concentration (ln Ci) with time (ti) produces a line with a slope equal to the negative inverse of the theoretical hydraulic residence time (HRT), Figure 1(a). Should tracer wash out of a digester at a rate faster than that predicted by the CSTR model, the presence of stagnant zones (or dead space) is indicated. Consequently, the mean hydraulic retention time (tm) obtained from the slope of the plot of ln Ci versus ti is less than the HRT, Figure 1(b). An RTD curve with a large peak near the origin, Figure 1(c), indicates that raw sludge is short-circuiting or bypassing from the inlet to the outlet of the digester. The plot of ln Ci versus ti, as a result, consists of two segments with different slopes. More complex departures from ideal mixing are analyzed by comparing observed tracer concentrations to predicted concentration from models developed by interconnecting idealized flow regions (plug flow, stagnant, and completely mixed) in different ways[5]. In analyzing RTD curves, the simplest mixing model should be used which fully predicts observed performance[6].

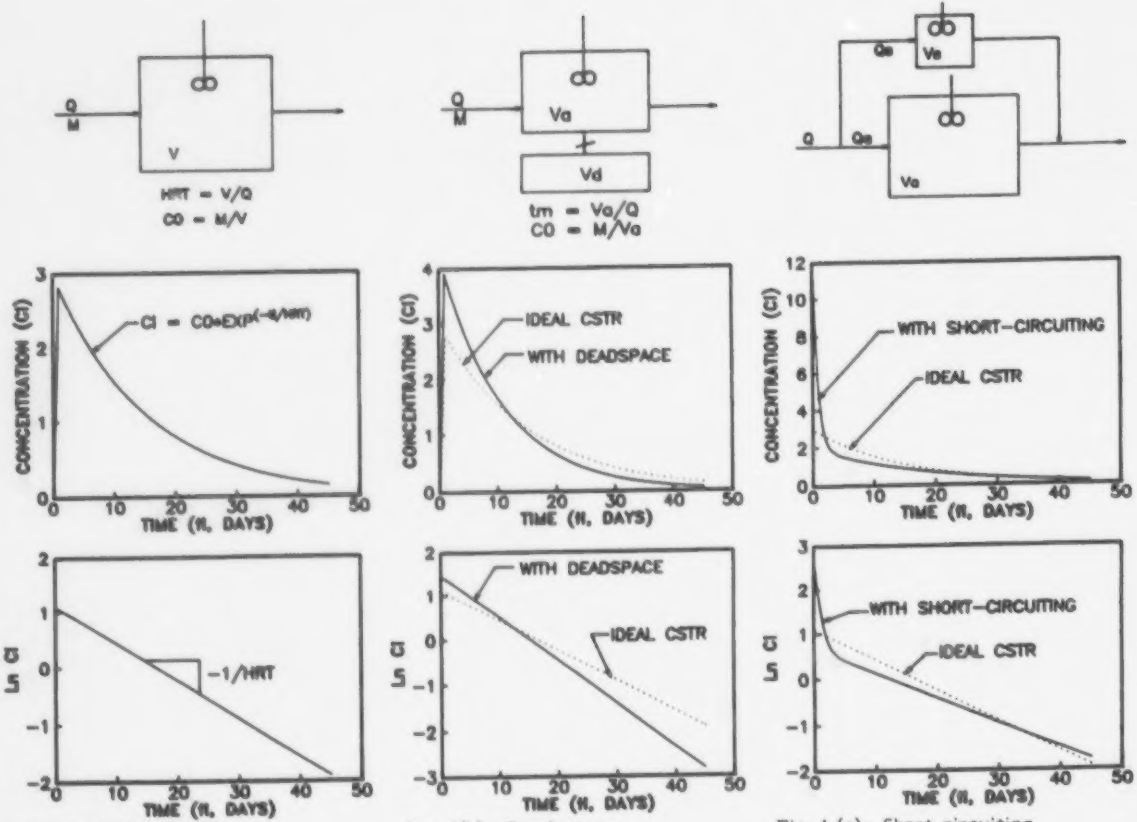

Fig. 1(a) - Ideal CSTR

Fig. 1(b) - Deadspace

Fig. 1 (c) - Short-circuiting

Fig. 1. Outlet concentrations of theoretical mixing models subjected to a pulse input of tracer

Results from tracer studies indicate that many high-rate digesters are poorly mixed. Smart[7] surveyed 10 digesters and found that four had short-circuiting. Estimates of active volume, that proportion of the total volume which can be considered perfectly mixed, averaged 46% percent for the 10 units. Monteith and Stephenson[8] estimated an active volume of 24% and a bypass flow of 61% at one high-rate digester; a second digester had an active zone of 23% without any detectable bypass. Such findings indicate that existing digester tankage is poorly utilized because of inadequate mixing.

To establish a relationship between mixer configuration and mixing performance, research was conducted at a primary high-rate anaerobic digester. The influence of the number of confined gas mixers, the applied mixing power, and the frequency of raw sludge pumping on the degree of mixing was established by generating residence time distribution curves and comparing the results to predictions from ideal mixing models.

Materials and Methods

Test Digester: The study was carried out at a primary digester at the Brantford, Ontario, sewage treatment plant. The digester had a diameter of 16.8 m and a sidewater depth of 6.4 m. Total digester volume was 1300 m^3. Postive displacement pumps transferred co-settled primary and waste activated sludge with a solids concentration of approximately 6% into the digester from the primary sedimentation tanks. Inlet piping discharged raw sludge at the top and the periphery of the tank (Figure 2). Digested sludge was withdrawn from the lowest point at the centre of the tank and transferred through a siphon overflow to the secondary digester. A 2.2 kW (3 HP) pump recirculated sludge through a spiral heat exchanger to maintain the contents of the digester at 35 deg. C.

Prior to the initiation of the research, the digester was mixed using 3 confined gas mixers and two 5.6 kW (7.5 HP) compressors. The gas mixers were located symmetrically 4.3 m from the centre of the digester, Figure 2. For this study, two additional 5.6 kW (7.5 HP) compressors and eight additional gas mixers were installed as shown in Figure 2. The inner three, or any combination of the outer eight mixers, could be placed in operation by opening or closing valves on the gas lines from the compressors.

Tracer Study Methodology: Three tracers--lithium chloride, tritiated water and sodium fluoride--were considered for the study. Because tritium is radioactive, approval could not be obtained from regulatory authorities for its use at a municipal treatment plant. Sodium fluoride was not selected because previous work indicated that variations in sludge solids concentration and ferrous sulphate originating from chemical addition for phosphorus removal interfered with the measurement of fluoride by the specific ion techniques[8,9]. In addition, interpretation of results is difficult for sodium fluoride because of high background fluoride concentrations resulting from the fluoridation of drinking water[10]. Accordingly, lithium chloride was selected.

Table 1 summarizes the mixing configurations for the six runs conducted at the Brantford digester. Runs were selected so that performance was evaluated over a range which included both the maximum and minimum combination of mixers and compressors. For runs no. 5 and no. 6, the same number of mixers and compressors were used but the location of the operating mixers was changed. Between runs no. 2 and no. 3, equipment was installed at the plant to control raw sludge pumping automatically, thereby changing pumping frequencies from 4 to 12 times per day. Mixing configurations were changed a minimum of 30 days prior to tracer injection.

At the beginning of each of the six runs, tracer was injected by dissolving 24 kg of technical grade lithium chloride in tap water and adding the solution to the raw sludge line through the suction side of the positive displacement pumps at the start of a raw

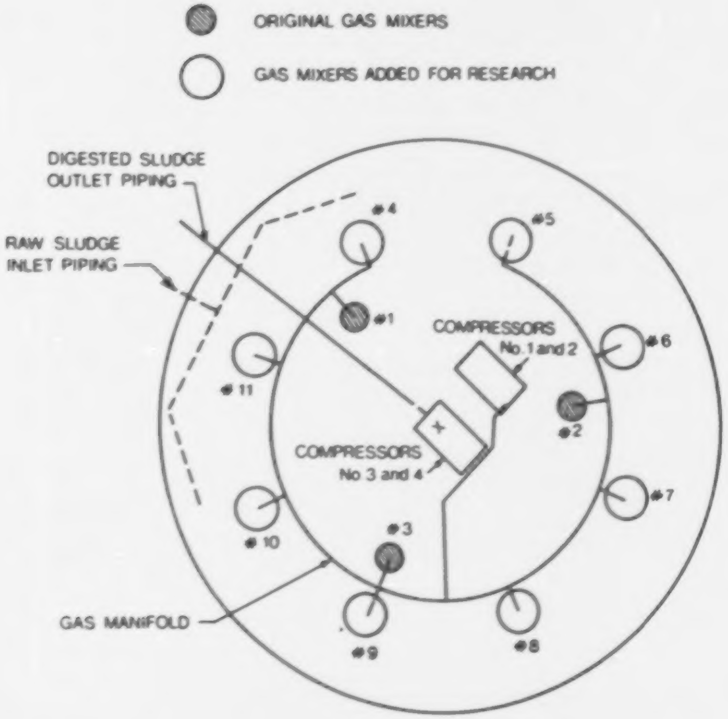

Fig. 2. Plan of Brantford digester showing location of gas mixers.

Table 1. Mixing Conditions for the Tracer Runs at the Brantford Digester

Run No.	Mixing Configuration	Mixer No.	Average Flow (m^3/day)	theoretical HRT (days)
1	1C:3N:4S	1-3	87.1	15.1
2	4C:8N:4S	4-8	137.0	9.6
3	1C:3N:12S	1-3	103.0	12.8
4	4C:8N:12S	4-8	120.7	10.9
5	2C:4N:12S	5,6,9,10	96.2	13.7
6	2C:4N:12S	7,8,11,4	96.2	13.7

Mixing configuration was designated as "C:N:S"--"C" for the number of compressors, "N" for the number of gas mixers, and "S" for the number of times raw sludge was pumped into the digester per day.

sludge pumping cycle. Assuming that the tracer was uniformly distributed throughout the contents of the digester, the expected initial lithium concentration (Co) for each run was 2.98 mg/L. Raw sludge flow rates were measured by recording pumping times after calibrating the sludge pumps from drawdown of the primary settlers.

Following tracer injection, samples of digested sludge were collected from a tap on the siphon overflow line between the primary and secondary digesters. Approximately 30 mL of digested sludge were collected in 60 mL polyethylene bottles labelled with the location, date, and time. To detect any short-circuiting of raw sludge, samples were collected once every 30 minutes for the first eight hours following injection. Subsequently, samples were collected once a day for a minimum of 2 to 3 HRTs (30 to 45 days for the Brantford digester). For some runs, the sampling frequency was once every other day.

Lithium Analysis: In preparation for analysis, 5 mL of concentrated nitric acid were added to the sludge sample to release lithium from the solid to the liquid phase. The digested sample was placed in a beaker and the plastic bottle rinsed with a 10% HCl solution to release lithium which had adhered to the sample bottle. The solution was boiled down to 10-15 mL and filtered through a Whatman No. 41 or 54 filter paper. The beaker was rinsed with 10% HCl. The resulting solution, with a volume of approximately 60 mL, was analyzed for lithium using flame atomic absorption.

Results

Tracer Concentrations and Recoveries: Figure 3 displays scatterplots of the lithium concentrations for the six runs. Excluded from the plots are the concentrations of samples collected during the eight-hour period immediately following injection. As these initial values varied widely from run to run, including them in the plots would have required the use of different scales and made visual comparison of the curves more difficult. The number above the corner of each scatterplot in Figure 3 is the maximum lithium concentration for the initial eight-hour sample period. A smoothed curve, based on a spline function and generated by computer[11], is shown in each plot to identify the general trend in the data. Points not conforming to the general trend were judged to be outliers and are identified in Figure 3 by an arrow. Outliers were deleted during data analysis.

Table 2 contains tracer recoveries, estimated by comparing the total mass of tracer recovered to the mass of tracer injected (3.93 kg Li). The product of the average raw sludge pumping rate over the length of run (Table 1) and the total area under the washout curve determined the mass of tracer recovered.

A number of preliminary conclusions can be drawn from the scatterplots in Figure 3 and the estimated tracer recoveries. For run no. 4, the curve is characteristic of a well mixed reactor, while for runs no. 1, no. 2 and no. 3 the curves are characteristic of poorly mixed conditions. Runs no. 2 and no. 5 display an intermediate degree of mixing. For run no. 3, tracer recovery was 48% and the lithium concentrations were lower than the corresponding concentrations for other runs. This indicates that there were problems in conducting the run. Either, one-half of the intended mass of lithium chloride was injected into the digester or valves on the raw sludge line were left open and not all of the tracer entered the digester. Excluding run no. 3, tracer recovery averaged 85%.

Mixing Performance Estimates: Mixing performance was quantified in terms of percentages for flow short-circuiting and deadspace. The volume of deadspace (Vd) was defined as that portion of the total digester volume which was stagnant, representing essentially wasted space in the digester, Figure 1(b). Short-circuiting occurred when flow (Qs) passed from the inlet to the outlet without entering the main active zone of

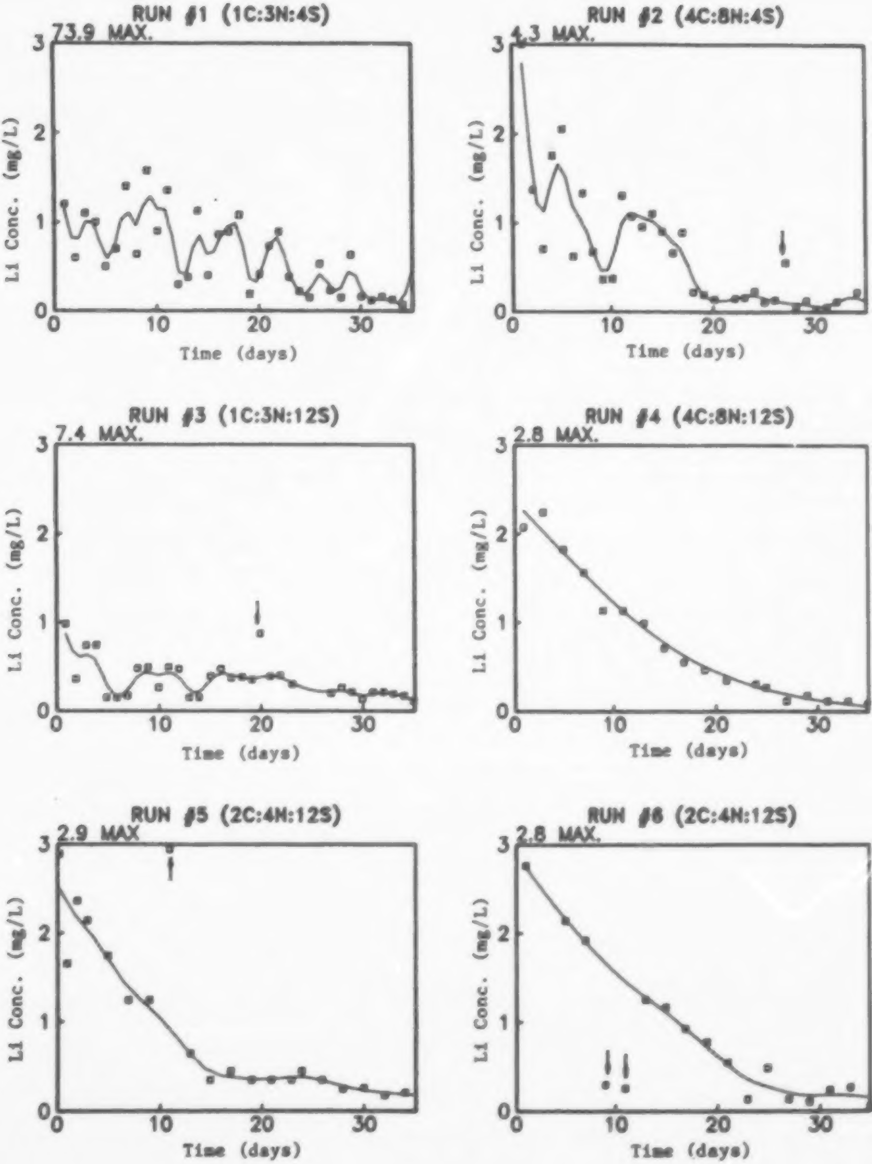

Fig 3. Scatterplots for tracer concentrations with smoothed curves to indicate trends.

Table 2. Tracer Recovery and Mixing Performance Estimates

Run No.	Recovery (%)	tm (days)	Qs/Q (%)	Cmax/Co	Deadspace (%)	Deadspace 95% C.I. (%)
1	83.2	10.8	28.8	29.5	49.1	30.6-59.8
2	79.3	8.1	0.4	1.4	16.0	< 0-30.7
3	47.8	13.3	11.6	4.9	--	--
4	89.3	9.8	0.0	0.9	10.1	2.4-16.6
5	74.0	10.5	0.0	1.0	23.4	7.8-34.4
6	89.0	12.6	0.0	0.9	8.0	< 0-34.0

mixing in the digester, Figure 1(c). For a reactor with short-circuiting and deadspace as defined by Figure 1(c), outlet concentrations (following pulse injection of tracer) are given by the following relationship[5]:

$$Cp = (Qa/Q)^2 * (Mr/Va) * e^{-t(Va/Qa)} + (Qs/Q)^2 * (Mr/Vs) * e^{-t/(Vs/Qs)} \text{ ....} (1)$$

Because short-circuiting occurs within a few hours following tracer injection, the above equation reduces to the following for sampling times greater than one day:

$$Cp = Co' * e^{-(t/tm)} \text{ ....} (2)$$

Estimates for the model parameters, tm and Co', for equation (2) were obtained by simple linear regression of the natural logs of the tracer concentrations with time. The slope of the regression equation was the negative inverse of the mean hydraulic residence time (tm). Values for tm are presented in Table 2.

Flow short-circuiting percentages were obtained for each run by partitioning the total area under the observed washout curve into two components as predicted by equation (2). Estimates of the area, As, resulting from tracer short-circuiting from the inlet to the outlet were obtained by applying the trapezoidal rule to paired values of Cs and t, where Cs was obtained from:

$$Cs = Ci - Cp \text{ ....} (3)$$

The ratio of the short-circuit area, As, to the total estimated area, At, under the washout curve provided an estimate of the short-circuiting flow fraction (Qs/Q). A discussion of the theoretical background for this approach and its verification using both simulated and experimental data can be found in Randle[12]. Table 2 summarizes the results obtained from the technique.

To provide a rough check for the short-circuiting percentages, ratios of the maximum measured tracer concentration, Cmax, to the expected initial concentration Co were calculated, Table 2. Co was 2.98 mg/L for all runs except run no. 3. Because the tracer recovery for run no. 3 was approximately 50%, a value of 1.49 mg/L was assumed. Values of Cmax/Co greater than one indicate short-circuiting. Both the Cmax/Co values and percentage short-circuiting estimates indicated that short-circuiting occurred for runs no. 1 to no. 3. In order of severity, run no. 1 displayed the greatest amount of short-circuiting and run no. 2 the least amount.

Deadspace values were based on the regression estimates of the mean hydraulic residence time of the actively mixed zone listed in Table 2. In cases where no short-circuiting was present, the percentage deadspace was calculated as:

$$\%V_d = 100\% * (1 - (t_m/HRT)) \quad \text{.......(4)}$$

In runs where short-circuiting was detected, the regression estimates for the mean hydraulic residence time represent the volume of the active zone divided by the flow into the digester. Accordingly, the active volumes were determined using the following relationship:

$$\%V_a = 100\%(1 - (t_m/HRT) * (1 - Q_s/Q) \quad \text{.......(5)}$$

The results are presented in Table 2. Because the estimate of the mean hydraulic residence was greater than the HRT for run no. 3, no estimate of deadspace is presented for this run in the table. The 95% confidence intervals for the deadspace estimates, shown in Table 2, were calculated based on the confidence interval of the slope obtained by simple linear regression of the natural logs of the tracer concentrations with time[13]. This approach assumes that the flow estimates are correct. The deadspace estimate for run no. 6 had a wide confidence interval because of the variability of a few observations at the tail of the washout curve.

Regression Estimate Checks: To check the validity of the regression estimates, the predicted concentrations, obtained from equation (2) using the regression estimates of the mean hydraulic retention time, were compared to the observed tracer concentration (Figure 4). Those observed concentrations judged to be outliers (Figure 3) were not included in the plots. In addition, for each run, values for the residual mean square were determined as follows:

$$MS_{res} = \frac{\Sigma (C_i - C_p)^2}{n - 2} \quad \text{.......(6)}$$

These values, displayed at the top left corner of the plots in Figure 4, quantify the fit of the observed to the predicted concentrations. For runs no. 4 and no. 6 which were well mixed, the plots and the low residual means square values demonstrate that the mixing model (equation (2)) was able to accurately predict the observed performance. In contrast, for runs no. 1 and no. 2, with a low degree of mixing, the overall fit between the observed and predicted values was poor. The mean square residuals for these two runs were roughly an order of magnitude greater than those for runs no. 4 and no. 6. In general, as the degree of mixing decreased, the fit between the observed and predicted values became poorer because the variability in the observed tracer concentrations increased. This finding brings into question the adequacy of the simple deadspace model (equation (2)) for quantifying mixing performance in the least mixed cases. The complexity of the tracer washout curves for the runs no. 1 and no. 2 suggested that sludge exchanged between a number of mixed and stagnant zones within the digester.

Other difficulties were encountered during data analysis. In contrast to RTD theory, the actual flow rates into the digester varied over the course of a run. Flow variability has the effect of increasing the variability of the tracer data in the outlet from the digester and introducing uncertainty into the HRT estimates used in conjunction with regression estimates of hydraulic residence time (tm) to estimate deadspace (equations (4) and (5)). In addition, the observed tracer concentrations which did not conform to the general shape of the washout curve were judged to be outliers and were deleted from subsequent analysis. The removal of one or two data points made a significant difference in the regression estimates for some runs. Similar problems were also encountered during regression in determining how much of a run to include in the data regression analysis; for some runs, regressing values over one HRT gave one estimate of mean hydraulic

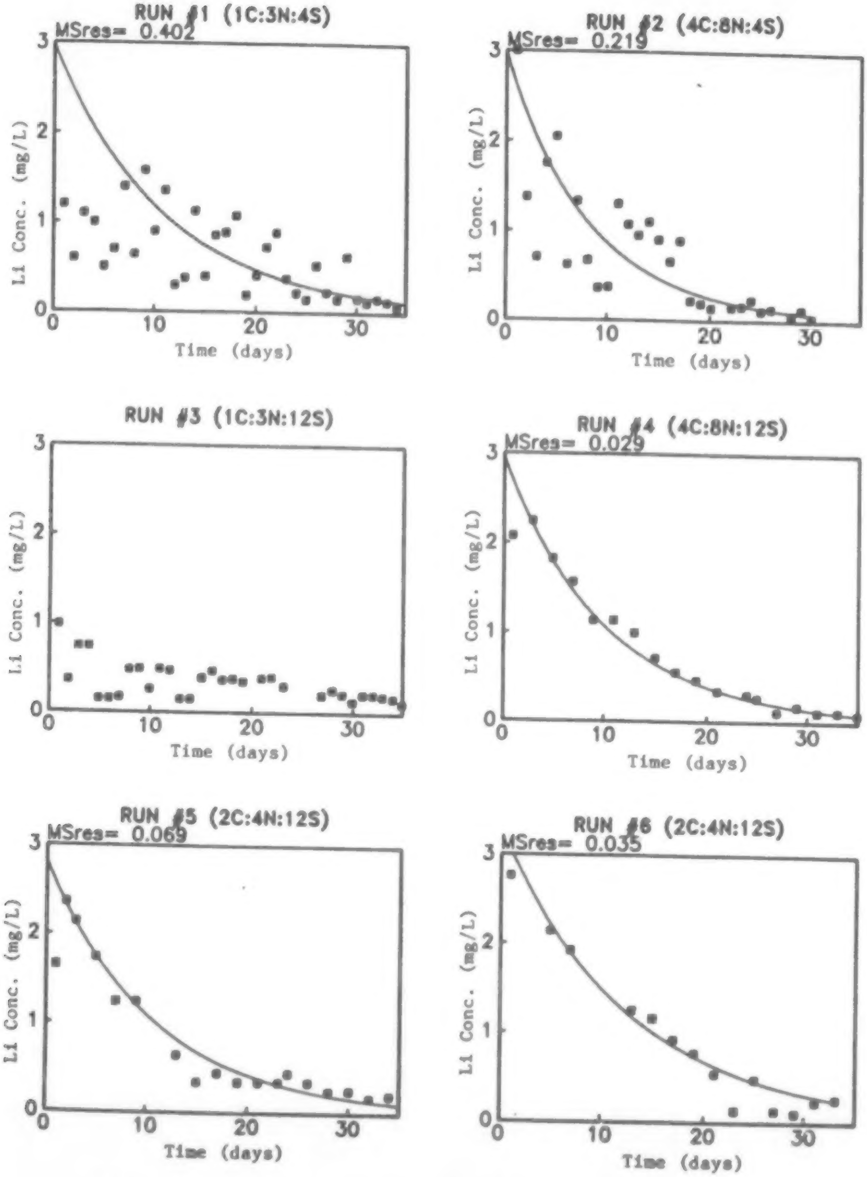

Fig 4. Scatterplots for tracer concentrations with predictions from regression models

residence time (tm) while regressing over 2 HRTs gave another. In this light, the mixing performance estimates contained in Table 2, particularly those for the poorly mixed runs, should be regarded as approximate.

Discussion

The results plotted in Figure 3 and contained in Table 2 demonstrate that changes in mixing configuration at the Brantford digester has a measurable effect on mixing performance as estimated by tracer response techniques. Direct conclusions could not be reached regarding the effect of improved mixing on digester performance as gas production and volatile solids destruction were not monitored as part of the study. However, the presence of short-circuiting in three of the six runs provided indirect evidence that digester performance was affected by mixing configuration. When short-circuiting occurred, a fraction of the raw sludge passed directly from the inlet to the outlet and therefore was not stabilized within the digester. Digester performance was accordingly reduced in proportion to the volume of raw sludge which short-circuited. To eliminate short-circuiting, designers must provide adequate separation between in the digester inlet and outlet. In addition, mixers should be positioned with respect to the inlet so that raw sludge enters the digester in a zone of active mixing. A major shortcoming of techniques such as vertical temperature or solids profiles, commonly used to measure digester mixing performance, is their inability to detect short-circuiting.

To further evaluate the results, unit power inputs were calculated and runs with the same sludge pumping frequencies (but different power inputs) were compared, Table 3. Values for these parameters increase as mixing performance improves, making any relationship between mixing performance and increased mixing power more apparent. Values for percent active flow and active volume were determined by subtracting the percent deadspace and short-circuiting estimates from 100%. Differences between active volume were listed in Table 3 if they were significant (based on the 95% confidence intervals in Table 2). Results for run no. 3 were not included because of the very wide confidence interval on the active volume.

Table 3. Comparison of Results at the Same Sludge Pumping Frequencies and Different Power Inputs

Run No.	Unit Power kW/1000 m^3	S	Qa/Q (%)	change	Va/V (%)	change
2	17.0	4	99.6	28.4	84.0	33.1
1	4.3	4	71.2		50.9	
4	17.0	12	100.0	0	89.9	13.2
5	8.5	12	100.0	11.6	76.7	NA
3	4.3	12	88.4		NA	

The results in Table 3 show that, as expected, both the active flow and the active volume estimates increased as mixing power increased. The greatest improvement resulted when power input was increased at the low sludge pumping frequency. In addition, the results summarized in Table 3 indicate that current design criteria for sizing digester mixing equipment may be inadequate. In all cases, active volumes were less than 100% of the total available digester volumes. For run no. 2 with a low sludge pumping frequency and a power input of 17.0 kW/100 m^3, the active volume was roughly 85% and there was a small but detectable flow fraction which short-circuited. Therefore, the contents of the Brantford digester were incompletely mixed at power inputs which exceeded US EPA recommendations[1] (5.3-7.9 kW/1000 m^3).

Further evidence for the inadequacy of current mixing criteria is provided by comparing unit power inputs for anaerobic digesters (5.3-7.9 kW/1000 m^3) to recommended values for other biological treatment processes. For instance, for aerobic digesters, Metcalf and Eddy Inc.[14] recommend that mixing powers in the range of 20-40 kW/1000 m^3 be applied. For aerated lagoons, research[15] indicates that complete mixing is not achieved below power inputs of 5.3 kW/1000 m^3. In complete contrast, however, Bruce[16] and Garber[17] claim that adequate self-mixing can be achieved from action of rising gas bubbles within the digesting sludge if the digester is loaded at a high enough rate.

To evaluate the effect of sludge pumping frequency, mixing performance estimates from runs with the same unit power inputs (but different frequencies) were compared (Table 4). Comparisons of run no. 1 with run no. 3 and run no. 2 with run no. 4 in Table 4 show that at two different power inputs the percent active flow increased with more frequent sludge pumping. The greatest increase in active flow resulted from increased pumping at the lowest power input. Even at the highest power inputs (run no. 2 and run no. 4), a slight but discernible improvement in active flow was achieved by increasing the pumping frequency. Therefore, the Brantford results suggest that digester mixing can be improved in a simple and cost-effective manner by pumping raw sludge into the digester at a rate which is as uniform and continuous as possible.

Table 4. Comparison of Results at the Same Power Inputs and Different Sludge Pumping Frequencies

Run No.	Unit Power kW/1000 m^3	S	Qa/Q (%)	change	Va/V (%)	change
3	4.3	12	88.4	17.2	NA	NA
1	4.3	4	71.2		50.9	
4	17.0	12	100.0	0.4	89.9	NS*
2	17.0	4	99.6		84.0	

NS*: the difference in active volumes for Runs no. 2 and no. 4 was not significant.

The published literature supports the finding that pumping frequency influences digester mixing performance[18]. More energy is required to disperse sludge which has been added as a slug than to disperse sludge fed into the digester at a uniform rate[19]. As a result, in chemical process industries, batch mixing systems are designed with 10-25% more power in comparison to continuous flow systems[20]. In addition, continuous feeding avoids problems created by slug loading of micro-organisms[1].

A number of factors in addition to power input, sludge pumping frequency, and mixer location influence the degree of mixing achieved by a digester. The fluid properties of the digesting sludge are complex (both pseudoplastic and thixotropic) and vary from plant to plant and over time. Frost[21] reports that the apparent viscosities of sludges from different plants varied by a factor of 50. A variety of mixer systems including confined and unconfined gas injection systems, mechanical stirring, and pumped recirculation are used at digesters[4]. In addition tank size and shape, the location of inlet and outlet with respect to each other and to the mixers, and the frequency of cleaning to remove accumulated grit influence mixing performance. Therefore, power requirements for anaerobic digester mixing systems are site specific and caution must be used in when comparison results from one digester with those from another.

Summary and Conclusions

Six residence time distribution curves were generated and analyzed to investigate the influence of mixing and the frequency of raw sludge pumping on the degree of mixing at a primary high-rate anaerobic digester. The principal findings of the study were as follows:

1) The presence of short-circuiting in three of the six runs provided evidence that digester performance was affected by mixing configuration. The prevention of short-circuiting in digesters is essential for maximizing digester performance.

2) Mixing performance (as estimated by flow short-circuiting and active volume fractions) improved following the installation of automatic equipment which increased the frequency at which raw sludge was pumped into the digester. The greatest improvement in mixing performance resulted from increased pumping frequency occurred at the lowest mixing power input.

3) Mixing performance improved following increases in total mixing power. The greatest improvement in active flow resulted when power input was increased at the low sludge pumping frequency.

4) Current power input criteria for digesters appear to be inadequate. The contents of the digester were incompletely mixed at power inputs which exceeded the US EPA recommendations (5.3-7.9 kW/1000 m^3). In addition, the literature indicates that unit power inputs for digesters are low in comparison to other biological wastewater treatment processes.

5) As the degree of mixing decreased, the fit between the observed tracer concentrations and those predicted values by a simple deadspace (equation (3)) model became poorer because the variability in the observed tracer concentrations increased. The complexity of the tracer washout curves for the least mixed cases suggested that sludge exchanged between a number of mixed and stagnant zones within the digester during the runs.

6) In addition to power input and pumping frequency, the literature reveals that mixing performance is influenced by the fluid properties of digesting sludge, tank configuration, mixer configuration and type and operating practices. Because these factors vary considerably from plant to plant, power requirements for anaerobic digester mixing systems are site specific.

Acknowledgements

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Notation

The following symbols are used in this paper:

As = area under RTD curve resulting from flow short-circuiting, mg/L*days;

At = total area under RTD curve, mg/L*days;

C = the number of compressors used per experimental run at the Brantford digester, dimensionless;

Ci = measured outlet tracer concentration, mg/L;

Cmax = the maximum measured tracer concentration, mg/L;

Cp = predicted tracer concentration from a mixing model, mg/L;

Cs = tracer concentration arising from short-circuiting, mg/L;

Co = the expected initial tracer concentration immediately following tracer injection, mg/L;

Co' = the initial tracer concentration of the actively mixed zone, mg/L;

M = the mass of tracer injected into the digester, kg;

Mr = the mass of tracer recovered during tracer study, kg;

n = number of samples, dimensionless;

N = the number of gas mixers, dimensionless;

MSres = the residual mean square (from predicted and observed tracer concentrations), $(mg/L)^2$;

Q = digester raw sludge flow rate, m^3/day;

Qa = the raw sludge flow rate entering the actively mixed zone in the digester, m^3/day;

Qs = the raw sludge flow rate short-circuiting or bypassing from the digester inlet to the outlet, m^3/day;

HRT = theoretical hydraulic residence time of the digester, days;

S = the number of times raw sludge was pumped into the digester during an experimental run at the Brantford digester, day^{-1};

t = elapsed time since tracer injection, days;

ti = the time, after injection, at which an outlet sample was collected, days;

tm = the mean hydraulic residence time, days;

V = the total digester volume, m^3;

Va = the digester volume which is actively mixed; m^3;

Vd = digester volume which is deadspace, m^3;

Vs = short-circuit or bypass digester volume; m^3.

DESIGNING ACTIVATED SLUDGE PLANTS FOR ENHANCED SUSPENDED SOLIDS REMOVAL

by

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Introduction

Activated sludge treatment plants produce a purified effluent by converting the soluble organics in a wastewater to biomass, entrapping colloidal particles in the biomass, and separating the biomass from the liquid stream. The conversion of soluble organics is very efficient, with most activated sludge plants producing effluents with a soluble BOD_5 of less than 5 mg/L (Benefield and Randall, 1980).

It is the escape of suspended solids from the final settler which limits the quality of the effluent achieved by an activated sludge plant. Most organic matter in the effluent is in the form of suspended solids. Depending on the hydraulic and solids retention time of an activated sludge plant, suspended solids which escape from the final settler will account for 50% to 90% of the effluent BOD_5 (Chapman, 1984).

The efficiency of phosphorus removal is also decreased by poor suspended solids removal. A study of the alum precipitation of phosphorus at an activated sludge plant indicated that total effluent phosphorus was highly correlated to the concentration of effluent suspended solids (U.S. EPA, 1971). Approximately 80% of effluent phosphorus was associated with the effluent solids.

Enhanced removal of suspended solids is achieved by careful attention to the design of three components of the activated sludge system--the pumps and controls governing the rate of inflow and recycle, the aeration tank and oxygen transfer system, and the final settlers. Pumps are selected and controlled to prevent the generation of hydraulic transients which can degrade effluent quality. The oxygen transfer system and the aeration tank are designed to provide a mixed liquor with good settling and clarification properties.

Initially, the final settler is designed to prevent a thickening failure by specifying the correct combination of final settler area and recycle rate for the particular settling characteristics of the mixed liquor. A potential for thickening failure exists if the rate at which solids enter the final settler exceeds the rate at which they are removed. The persistence of such an imbalance will cause solids to be added to the sludge blanket, eventually resulting in a massive loss of solids when the blanket reaches the weirs of the settler. Techniques developed by Dick (1970) and Keinath et al. (1979) enable the designer to assess the thickening capacity of a final settler.

Having ensured that the final settler has adequate thickening capacity, consideration is given to optimizing the clarification function of the unit. Clarification is controlled by overflow rate, tank depth, and the mixed liquor suspended solids (MLSS) concentration. Recent designs have sought to maximize clarification by the use of flocculating centrewells and effluent weirs with horizontal baffles to deflect density currents.

This paper describes the mechanisms governing the removal of suspended solids by the activated sludge system. Various aspects of the design of influent and recycle pumps, the aeration tank and oxygen transfer system, and final settler are discussed as they impact on the removal of suspended solids. The material contained in the paper is based on a review of the current literature and on studies conducted by the author (Chapman, 1983, 1984, 1985).

Pump Selection and Control

Hydraulic transients or surges in flow rate can be generated by influent or recycle pumps. Because activated sludge plants do not completely attenuate hydraulic transients and final settlers respond to sudden changes in hydraulic loading, pump-induced transients degrade effluent quality. To avoid propagating transients, pumps and pump control schemes must be selected with care. In the aeration tank, outlet weirs can be designed to enhance hydraulic dampening.

Hydraulic Transients

The technical literature documents the occurrence of substantial flow variations originating from the operation of influent and recycle pumps. For example, Maskell and Lumbers (1974) in England found that the action of lift stations combined to produce an instantaneous influent flow rate which cycled every 50 minutes from a low of zero to a high which was more than twice the dry weather capacity of the plant. Figure 1, from Olsson et al. (1984), is a plot of the influent flow rate at the activated sludge plant in Kirkeskoven, Denmark. The action of the influent pumps in producing high frequency/high amplitude transients is evident.

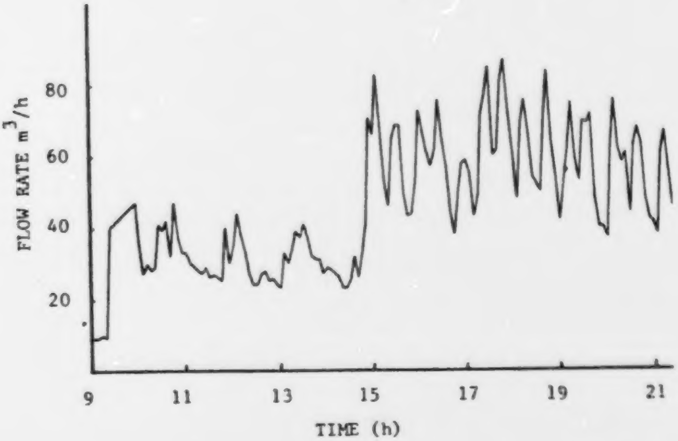

Fig. 1 Influent flow rate pattern for an activated sludge plant showing pump-induced hydraulic transients (after Olsson, et al., 1984)

At a plant in Florida, Collins and Crosby (1980) observed flow rate variations resulting from the intermittent operation of recycle pumps. The results, presented in Figure 2, show a 2.5 fold change in effluent flow rate over the course of a cycle. As will be shown, variations in hydraulic loading of this magnitude significantly deteriorate effluent quality.

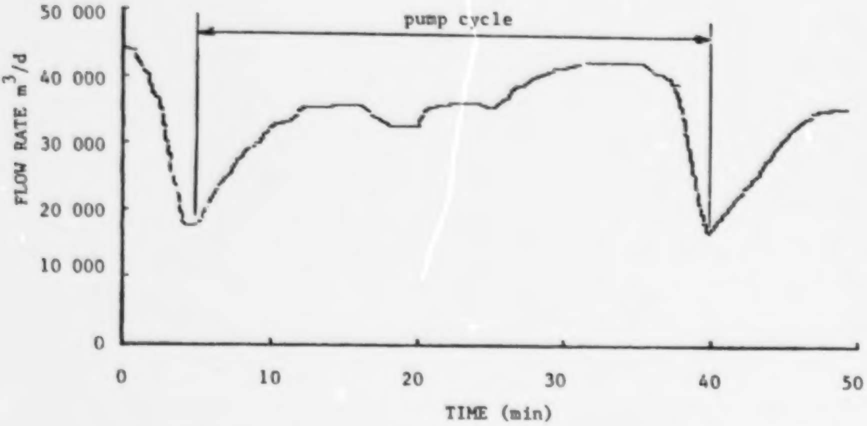

Fig. 2. Final effluent flow rate showing effects of intermittent sludge recycle (after Collins and Crosby, 1980).

Hydraulic Dampening in Conventional Activated Sludge Plants

Conventional activated sludge designs have limited capacity to dampen pump-induced transients (Bender and Crosby, 1984). The ability of a tank to dampen flow transients is determined by the relationship between the flow discharged by the tank outlet and the liquid level in the tank. Four factors govern the relationship between flow and liquid level--weir type, the number of weirs in the outlet, the dimensions of the weirs, and the surface area of the tank (Olsson et al., 1984). Because activated sludge plant use large numbers of V-notch weirs in the outlets, Andrews (1974) pointed out most plants are essentially constant volume systems. With V-notch weirs, the flow rate is proportional to liquid level as follows:

$$Q = kh^{2.5}$$

where: Q = flow
k = a constant (consisting of the product of the weir coefficient and the number of weirs)
h = head over the weir.

Therefore, a large increase in flow rate results from a small change in liquid level. Hill (1985) determined that the hydraulic time constant of a conventionally designed activated sludge plant was approximately five minutes.

Final Settler Dynamics

Flow disturbances which enter the final settler degrade effluent quality. Experimental evidence has established that the dynamic response of final settlers to sudden changes in flow rate is nonlinear (Chapman, 1985; Olsson and Chapman 1985; Hill, 1985; Dietz and Keinath, 1984). Sudden increases in flow rate result in rapid increases in the concentration of effluent suspended solids; sudden decreases in flow rate result in a more gradual decay in effluent solids. Such findings support the contention of Crosby and Bender (1980) that turbulence within a final settler is generated quickly but decays slowly.

Preventing and Attenuating Transients

Frequently, influent pumps are sized to handle wet weather flows and turned on or off by float switches in the wet well of the pump station. During periods of dry weather flow, the operation of these pumps can result in substantial hydraulic transients. To avoid this, mechanical or electrical variable speed drives can be used in conjunction with influent pumps. The rate of influent pumping can be then be changed smoothly from one level to another. As an alternative to variable speed drives, a series of single- or dual-speed pumps of different sizes can be selected and scheduled to operated to avoid large incremental changes in flow rate. For small treatment plants, consideration can be given to providing equalization.

Design features can be incorporated in the aeration tank to enhance the ability of the plant to dampen hydraulic transients once they have been generated. Increasing both the freeboard of the aeration tank and the hydraulic resistance of the outlet weir in the tank will increase the degree of dampening. By manipulating the number, type, and dimensions of the weirs, the designer can achieve the desired result. Reference should be made to a paper by Olsson et al. (1984) for additional details and discussion of hydraulic transients and aeration tank weir design.

Substantial benefits can accrue by avoiding the propagation of pump-induced transients. The City of Detroit, for instance, instituted surge control of influent pumping thereby eliminating the need for four new settler and saving $U.S. 18 million (Porta et al., 1980).

Aeration Tank Design

The activated sludge process employs a mixed culture of microorganisms to remove soluble organics and entrap colloidal particles. Individual bacteria are themselves colloidal and as such cannot be removed by sedimentation. Therefore, effective treatment by the activated sludge system depends critically on aggregating individual cells into flocs which can be removed by sedimentation. The degree of bacterial flocculation and the settling properties of the flocs are determined by growth conditions within the aeration tank. Solids retention time (SRT) or sludge age, the level of dissolved oxygen (D.O.), the intensity of turbulence within the tank, and the tank configuration determine how well a mixed liquor grown in the aeration tank will thicken and clarify in the final settler.

The Nature of Activated Sludge Flocs

Flocs are packets of negatively charged bacteria consisting of both filamentous (or branch-like) bacteria and floc forming (or spherical) bacteria (Palm et al., 1978). The presence of filaments provides a "backbone" which increases the resilience of the floc. However, under certain conditions--low concentrations of substrate, nutrients, or dissolved oxygen--filamentous organisms have an advantage over floc forming organisms and grow at a faster rate. Filaments then grow out of the floc into the surrounding liquid (Figure 3), producing a biomass with a low specific gravity which settles slowly. This condition, termed filamentous bulking, increases the potential for thickening failure in the final settler.

The "glue" which holds individual bacteria together to form flocs consists of long-chained molecules called exopolymers (Tenney and Verhoff, 1973). These polymers are produced by the bacteria and consist mainly of polysaccharides (Brown and Lester, 1980). Protozoa in the mixed liquor may also contribute to the process of bacterial floc formation by feeding on individual cells or producing a type of biopolymer. However, research with synthetic waste has indicated that the flocculation of bacteria can occur without the presence of protozoa (Hughes and Stafford, 1976).

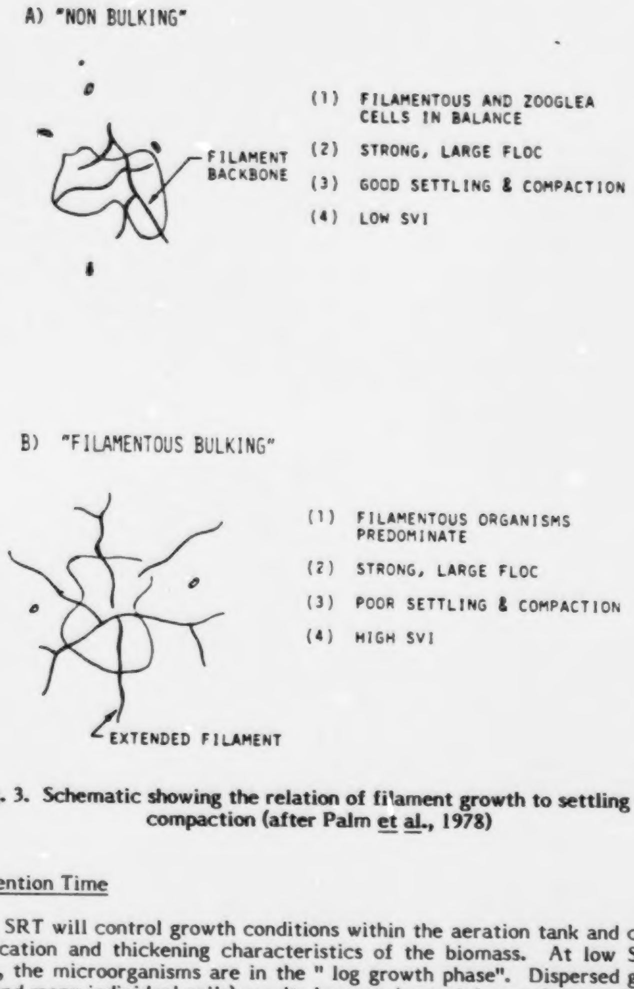
A) "NON BULKING"
FILAMENT BACKBONE
(1) FILAMENTOUS AND ZOOGLEA CELLS IN BALANCE
(2) STRONG, LARGE FLOC
(3) GOOD SETTLING & COMPACTION
(4) LOW SVI
B) "FILAMENTOUS BULKING"
(1) FILAMENTOUS ORGANISMS PREDOMINATE
(2) STRONG, LARGE FLOC
(3) POOR SETTLING & COMPACTION
(4) HIGH SVI
EXTENDED FILAMENT

Fig. 3. Schematic showing the relation of filament growth to settling and compaction (after Palm et al., 1978)

Solids Retention Time

Control of SRT will control growth conditions within the aeration tank and consequently the clarification and thickening characteristics of the biomass. At low SRTs, below three days, the microorganisms are in the " log growth phase". Dispersed growth (with few flocs and many individual cells) results because bacterial growth exceeds biopolymer production. Conversely, at high SRTs, greater than 15 days, excess bacterial polymer restabilizes biocolloids, producing pin point floc.

There are other problems associated with long SRTs. First, they increase the biomass concentration in the aeration tank, increasing the solids loading to the final settler. Second, at long SRTs, there is a tendency for nitrifying bacteria to become established. If nitrification occurs in the aeration tank, then sludge will tend to denitrify in the final settler. During denitrification, nitrogen and carbon dioxide gas bubbles become trapped in the flocs causing some of them to float (Crabtree, 1983). Floating solids add to the concentration of suspended solids discharged in the final effluent. Because of the problems associated at long SRTs, hydraulic loading rates are decreased for final settlers at extended aeration plants.

Dissolved Oxygen

Oxygen transfer equipment must be selected and installed to provide adequate but not excessive levels of D.O. If the levels of D.O. are too low in the aeration tank, then filamentous bulking may result with the attendant possibility of thickening failure. It is current practice to maintain a minimum concentration of 2 mg/L in the aeration tank. The environment in aeration tanks is inherently disruptive to activated sludge flocs (Parker et al., 1972). D.O. levels in excess of those required for microbial growth and the prevention of bulking, unnecessarily increase the level of turbulence within the aeration tank and strip colloidal particles from the flocs.

To minimize floc shear and avoid filamentous bulking, D.O. levels can be maintained within selected limits by D.O. sensors and control equipment for the blowers and the air distribution piping. The use of pure oxygen systems or fine bubble diffusers ensures that adequate levels of D.O. are provided while minimizing the intensity of turbulence within the aeration tank. Tapered aeration also minimizes floc shear. Because the respiration rate of the biomass is low at the discharge end of the aeration tank, less air is supplied to this region with tapered aeration. Turbulence is therefore reduced and fewer nonsettleable particles are stripped from the flocs.

Tank Configuration

The configuration of the aeration tank can play a role in determining the characteristics of the biomass. For example, the provision of a substrate gradient along the length of the tank favours the growth of floc forming bacteria and decreases the potential for filamentous bulking. Tapered aeration and pure oxygen systems both provide substrate gradients. To reduce the potential for floating solids arising from denitrification in the final settler, the aeration tank can be modified to include a small anoxic zone at the inlet end of the aeration tank. Experience in England indicates that, with a retention time of 30 to 40 minutes, the zone consistently achieves 50% denitrification (Briggs and Jones, 1978).

Final Settler Design

The momentum of the flow of mixed liquor entering the final settler generates turbulence within the unit. Because of this turbulence and the flocculent nature of the solids, overflow rate cannot be used as the sole criteria to size final settlers. Tank depth and MLSS concentration also influence the removal of suspended solids in the final settler. To promote enhanced removal of suspended solids, final settlers are being designed with flocculating centrewells and specialized baffles.

Solids Removal in Final Settlers

A final settler contains two zones defined by the top of the sludge blanket. Above the blanket is a turbulent zone of clarified liquid (supernatant). Within the blanket, solids separated from the supernatant undergo thickening and compression. Separation is accomplished by the mechanisms of flocculation and sedimentation above the interface and the mechanism of entrapment at the interface as the incoming mixed liquor impinges on the blanket (Gregory, 1979).

The flow of mixed liquor into the final settler generates turbulence. For a circular settler with a centre feedwell, the incoming flow plunges downwards, flows across the top of the sludge blanket, and then upwards along the walls of the settler (Figure 4). A portion of the flow is directed back towards the feedwell (Anderson, 1945). The circular roll in the tank is a product of the momentum of the flow entering the final settler and not exclusively, as frequently stated, by density differences between the mixed liquor and the supernatant (Murphy, 1964). The circular roll creates both mixing within the final settler and the potential for resuspending solids.

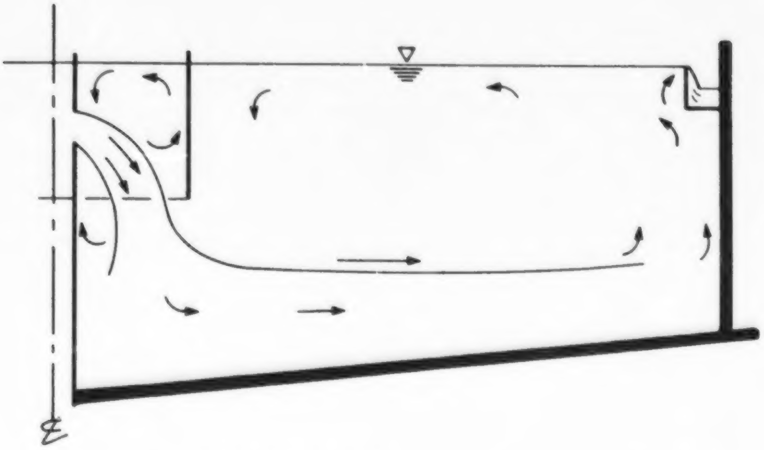

Fig. 4. Flow pattern in circular settler created by incombing mixed liquor (after Tekippe, 1984)

Overflow Rate

Overflow rate or hydraulic loading rate is defined as the plant inflow rate per unit of surface area of the final settler. Based on the work of Camp (1953), overflow rate has been used by the engineering profession as the fundamental parameter for designing final settlers. Camp demonstrated that the hydraulic overflow rate for a tank handling discrete particles is equal to the sedimentation velocity of the slowest settling particle which is completely removed.

U.S. EPA (1975) design criteria for average overflow rates are in the 16 to 32 m^3/m^2/day range for conventional activated sludge plants. These recommended values are decreased for extended aeration plants and increased for pure oxygen plants because of the previously discussed influence of these processes on the settling properties of the mixed liquor. In recent years, there has been a trend towards the use of more conservative average overflow rates for conventional activated sludge plants, from 12 to 16 m^3/m^2/day (Tekippe, 1984).

Two cautionary points should be made with respect to the use of overflow rate in designing final settlers. First, in some cases, it has been used to the exclusion of the other design parameters such as depth (Mazurczyk et al., 1980). Camp's analysis dealt with discrete particles. Final settlers remove particles which are flocculent in nature. Therefore, such criteria as depth (discussed in the next section) become important.

Secondly, the definition of overflow rate tends to convey the impression that the fluid within the final settler rises at a uniform velocity across the entire cross section of the tank. It does not. As has been discussed, most circular sedimentation tanks are subject to "density currents" which create a circular motion and nonuniform velocities. Therefore, although overflow rate is an important design parameter, its use should not preclude the use of other design parameters nor should its use be taken as implying that the fluid rises at a uniform rate across the entire cross-section of the tank.

Tank Depth

Because mixed liquor is flocculent, depth is an important design parameter for final settlers. Increasing the depth of the final settler increases the time available for flocculation in the supernatant zone above the sludge blanket interface. Secondly, a deep tank provides storage for biosolids displaced from the aeration tank by storm flows. With a deep tank, the depth between the top of the sludge blanket and the effluent weir can be increased, decreasing the likelihood that solids scoured from the blanket will reach the weirs. Parker (1983) examined the influence of tank depth on effluent quality by collecting data from a number of settlers with similar overflow rates. He found that both the level and variability of the effluent suspended solids concentration decreased as the depth of the settler increased.

Traditionally, shallow tanks have been employed because of the perception that cost penalties are associated with deep tanks. Tekippe (1984) examined capital costs for two different tank designs, each providing a volume of 4400 m^3. One design called for a shallow tank (3.7 m deep) with a 39 m diameter; the second design employed a depth of 6.1 m and a diameter of 30 m. The cost estimate for the shallow tank was $74 000 more than that for the deep tank. The rake mechanism was the major source of cost difference between the designs. Because of its smaller diameter, the deep tank had a less expensive rake mechanism. Although the costs for tanks are site specific and depend on soil conditions, deep tanks should not be automatically ruled out as uneconomical.

Because of their advantages, deeper tanks are becoming more common. In the 1959 version of the ASCE/WPCF manual, recommended depths were 2.1 to 3.7 m. By contrast, the 1977 manual suggested a depth of 4.6 m for tanks with a diameter of 31 m (ASCE/WPCF, 1977). A survey of major U.S. consulting firms indicated that the average depth of new tanks was 4.6 m with a range from 3.7 to 6.1 m (Tekippe, 1984).

Mixed Liquor Suspended Solids Concentration

The MLSS concentration in the feed to a final settler influences effluent quality. Chapman (1983) found that effluent suspended solids increased at a rate of 4 mg/L for every g/L increase in MLSS.

MLSS is a key parameter in the optimization of overall capital costs for the activated sludge system. For low MLSS values, the required volume for the aeration basin is large while for the final settler it is small. For high MLSS values, the situation is reversed. Tuntoolavest et al. (1980) suggested that optimal activated sludge designs call for low MLSS concentrations.

For periods of hydraulic overloading, the step feed modification can improve settler performance by manipulating MLSS concentration. During periods of high flow, the step feed configuration enables the point of addition of the influent to be shifted towards the outlet end of the aeration basin. Consequently, biomass is stored at the inlet end of the aeration tank, while the flow entering the settler has a reduced MLSS concentration. The net effect is to reduce the solids loading to the settler and improve the effluent suspended solids concentration.

Special Features

Two novel features have been incorporated in some recent final settlers. The first is a special inlet device which promotes flocculation as the mixed liquor enters the final settler. The device consists of an influent diffusion well which is closed at the bottom and an outer flocculating centrewell (Figure 5). The mixed liquor flows between the two via gates which can be controlled either automatically or manually. The size of the gate openings can be adjusted to control the energy level for flocculation in the outer well. The dimensions of the inner well and the outer baffle are 6% and 25% respectively of the tank depth. Parker (1983) reported achieving suspended solids concentrations of less than 10 mg/L using floculator centrewells and tanks depths of 4.6 m.

A second novel feature being added to final settlers is a short horizontal baffle placed underneath the effluent weir (Figure 6). The baffle deflects the circular roll away from the effluent weir and prevents floating solids from entering the annular area between the scum baffle and weir (Stukenburg et al., 1983). By installing the baffle in a circular settler at Stanford, Connecticut, Crosby (1982) observed a reduction in effluent solids from 19 mg/L to 12 mg/L.

Summary

Suspended solids account for the majority of BOD_5 in the effluent from an activated sludge plant. Similarly, the efficiency of phosphorus removal by precipitation is reduced by the escape of suspended solids in the final effluent.

Poor suspended solids removal can be attributed to a number of factors. Hydraulic transients which are generated by influent or recycle pumps degrade effluent quality. In the aeration tank, inadequate attention to the selection of SRT, D.O. levels, and tank configuration can result in a floc which settles and clarifies poorly. A circular flow pattern in the final settler is generated by the momentum of the incoming mixed liquor and density differences between the mixed liquor and the clarified supernatant in the settler. This current creates both mixing and scour within the settler.

The removal of suspended solids by the activated sludge process can be enhanced by:

1. selecting and controlling influent and recycle pumps to avoid the creation of large hydraulic transients,
2. increasing the freeboard and the hydraulic resistance of the outlet weirs of the aeration tank to maximize dampening,
3. selecting an SRT with due regard to its effect on the settling and clarification properties of the mixed liquor,
4. preventing sludge bulking by providing adequate levels of dissolved oxygen in the aeration basin and selecting process configurations (such as tapered aeration) which create a substrate concentration gradient along the length of the tank,
5. reducing levels of turbulence within the aeration tank by selecting efficient oxygen transfer equipment and controlling the level of D.O.,
6. reducing floating solids in the final settler using a small anoxic zone in the aeration tank to provide denitrification prior to final settling or reducing the SRT to minimize nitrification,
7. reducing the MLSS concentrations by reducing the SRT, designing larger aeration tanks, or making provision for step feed,
8. designing final settlers with adequate tank depth, and
9. incorporating special features into the design of final settlers such as flocculating centrewells or effluent weirs with a short horizontal baffle to deflect the circular current created by the incoming flow.

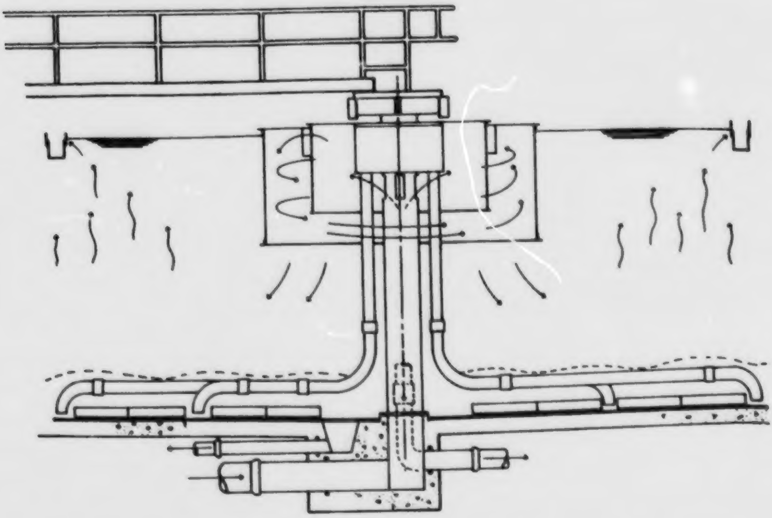

Fig. 5. Final settler with flocculating centrewell (after Parker, 1983)

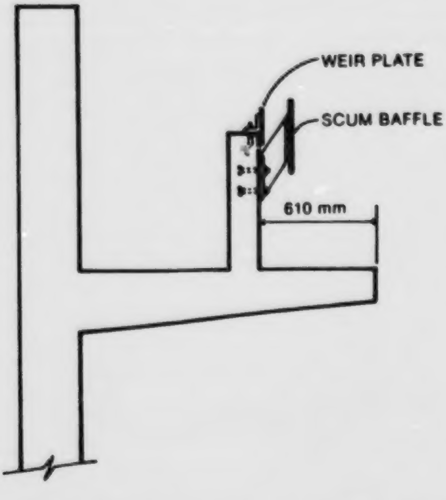

Fig. 6. Cross-section of weir through showing horizontal baffle to deflect circular current (after Stukenburg et al., 1983)

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WASTE METALS - THE CANADIAN APPROACH TO LIMITING METALS ON LAND FROM MUNICIPAL SLUDGES

by

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Abstract

A summary of Canadian information concerning the production and agricultural utilization of municipal sludge is presented. Sludge quantities (1981) are estimated and waste guidelines for agricultural utilization are compared. Research findings on waste metal uptake by agricultural crops are reviewed. The future of agricultural utilization is considered.

Introduction

Sewage sludge is a nutrient rich, largely organic by-product of municipal wastewater treatment. It must be removed from the treatment facility and the disposal options are ocean dumping, incineration, landfilling and utilization on agricultural land. Recent high energy costs and environmental awareness have increased interest in sludge disposal by utilization on agricultural land. Managed properly, this practice minimizes environmental risks, takes advantage of the fertilizer and soil conditioning value of sludge, and is frequently the least expensive disposal option.

There are two types of environmental risk (temporary and persistent) associated with sludge utilization on agricultural land. **Temporary risks** disappear within one year or at most a few years following sludge application and include malodour, pathogens, groundwater contamination with nitrate-nitrogen, and phytotoxicity due to soluble salts or toxic biodegradation products from inadequately stabilized sludge. **Persistent risks** remain long after the temporary ones have disappeared and include increased concentrations of industrially produced organic compounds such as polychlorinated biphenyls (PCB's) and waste metals in soil. It is estimated that the half-life of persistent organics such as PCB's in soil is about ten years (Fries, 1982) while that of most waste metals is about one thousand years (Bowen, 1975). "Waste metals" as used in this paper refers to metals and non-metals (e.g., As, B, F) of environmental concern.

It is widely accepted that sewage sludge utilization on agricultural land must be regulated to minimize the risks associated with heavy metal buildup in soil. Consequently, guidelines for this purpose have been developed in many countries including Canada. The objectives of this paper are to present Canadian information on:

(a) municipal sludge production and agricultural utilization,
(b) waste metal guidelines for agricultural utilization,
(c) research findings on waste metal uptake by agricultural crops, and
(d) future agricultural utilization.

Municipal Sludge Production and Agricultural Utilization

Based on 1981 data (Table 1), a majority of the Canadian population is served by sewers. The remainder, mainly rural, is served by septic tanks. Approximately one-half of the

Table 1. Canadian Population (National Inventory, 1981) **and Estimated Municipal Sludge Quantities for 1981**

Jurisdiction	Population (10^6 people)			Stabilized Sludge (10^3 tonnes dry wt)		
	Total	Sewered	Sewage Treatment	Total[a]	Disposal	Agricultural Utilization[c]
British Columbia	2.72	1.64	1.57	28	10	
Alberta	2.21	1.54	1.49	63[b]	3	40
Saskatchewan	0.96	0.57	0.57	15.4[b]	6.2	3.8
Manitoba	1.02	0.79	0.79	20[b]		17
Ontario	8.55	6.58	6.37	192[b]	125	61
Quebec	6.38	5.48	0.41	82	6.7	
New Brunswick	0.69	0.37	0.19	7.1	3.3	
Nova Scotia	0.84	0.45	0.11	7.5	2.5	
Prince Edward Island	0.12	0.06	0.05	0.9		
Newfoundland	0.56	0.32	0.06	5.1	1.1	
Yukon Territory	0.02	0.02	0.02	0.3		
Northwest Territories	0.04	0.04	0.02	0.7		
Canada	24.1	17.8	11.6	422	158	122

[a] Estimated for the sewered population assuming 40 and 64 g/capita·day of anaerobically digested sludge, respectively, for populations served by primary and activated sludge plants (Schmidtke, 1981), 50 g/capita·day for population served by lagoons and 40 g/capita·day assuming anaerobic digestion of primary sludge for sewered population with no sewage treatment.

[b] Adjusted to reflect increased sludge quantities resulting from tertiary treatment (e.g., chemical addition for phosphorus removal).

[c] Based on sludge removal from primary and waste activated plants only. Data for Alberta, Manitoba and Ontario are measurements whereas other data are estimates.

Canadian population is served by sewage treatment. In British Columbia, Alberta, Saskatchewan, Manitoba, Ontario, Prince Edward Island and the Yukon Territory, the populations served by sewage treatment approximate those served by sewers. However, in Quebec, New Brunswick, Nova Scotia, Newfoundland and the Northwest Territories, many fewer people are served by sewage treatment than by sewers. With the exception of Quebec and the Northwest Territories, these jurisdictions are maritime, and raw sewage is released to rivers and the Atlantic Ocean. In Quebec, raw sewage released to rivers and lakes has caused serious water pollution problems and a very large program of sewage treatment plant construction has been undertaken. In the Northwest Territories one-half of the sewage does not receive treatment.

Twenty-nine percent, of the estimated 1981 total sludge production in Canada, was utilized on agricultural land and thirty-seven percent was disposed of by landfilling, incineration and ocean dumping (Table 1). The remaining thirty-four percent represented sludge which would have resulted from sewage treatment (e.g., in Quebec) and sludge contained in sewage treatment lagoons, which is removed irregularly, if ever. However, this breakdown differed widely among jurisdictions within Canada. For example, most of the sewered population in Ontario was served by mechanical wastewater treatment, and agricultural utilization and disposal accounted for 97% of the sludge production. Agricultural utilization was practised widely in Ontario and accounted for 32% of the sludge production. By contrast, there was very little wastewater treatment in Quebec in 1981 and sludge disposal accounted for only 8% of production. Agricultural utilization of sludge in Quebec was prohibited by legislation. Large proportions of the sludge produced in Alberta, Saskatchewan and Manitoba were utilized on agricultural land. In Alberta, land applied sludge originated mainly from the cities of Calgary and Edmonton and in Manitoba from the city of Winnipeg. The Atlantic provinces and British Columbia had expressed interest in agricultural utilization, however, this practice did not account for significant quantities of sludge. In British Columbia there was considerable interest in composting sludge to produce horticultural potting media and soil conditioners.

Waste Metal Guidelines for Agricultural Utilization

Agricultural utilization of sludge in Canada comes under both federal and provincial jurisdiction. The sale of sludge and sludge-based products as fertilizers and soil amendments is within the purview of the Fertilizers Act and Regulations administered by Agricultural Canada, and a federal trade memorandum (Standish, 1981) has been developed to regulate this activity. However, utilization of sludge supplied free of charge is within provincial jurisdiction and there is variable regulation as follows. Ontario (Ontario, 1986), Alberta (McCoy et al., 1982) and Nova Scotia (Nova Scotia, 1983) have adopted guidelines; British Columbia (British Columbia, 1982), Saskatchewan (Saskatchewan, 1987) and Quebec (Quebec, 1986) have developed draft guidelines; Manitoba (Manitoba, 1986) has adopted a guideline for Winnipeg sludge utilization; and the remaining provinces have no guidelines. The waste metal guidelines for agricultural utilization of sludge in Canada vary with jurisdiction as indicated in Table 2.

Ontario. Waste metal guideline development for agricultural utilization of sludge in Canada originated in Ontario during the early 1970's following initiation of the phosphorus removal from wastewater program undertaken in the lower Great Lakes region. The phosphorus removal program resulted in both increased sludge production and increased pressure for sludge disposal by utilization on agricultural land.

Following a thorough review of the information available on sludge utilization in agriculture, background waste metal concentrations were determined for more than 200 Ontario agricultural soils (0-15 cm depth). Maximum acceptable concentrations in soil were defined assuming a doubling of the background levels and maximum acceptable loadings as follows: As, 14; Cd, 1.6; Co, 10; Cr, 30; Cu, 50; Hg, 0.2; Mo, 4; Ni, 32; Pb, 30; Se, 0.8; Hg, 0.2 and Zn, 110 kg/ha were calculated. As further experimental information and experience were obtained, the limits for Co, Cr, Hg, Pb, Se and Zn were increased to their present levels (Table 2).

Liquid anaerobically digested sludge is the principal form of sludge available for land application in Ontario. However, maximum acceptable metal concentrations were not defined for these sludges. They contain appreciable amounts of nitrogen and it was recommended that their application rates be based upon plant-available nitrogen concentration. Thus, ammonium plus nitrate-nitrogen to heavy metal ratios were defined to determine acceptability for use in agriculture (Ontario, 1986).

By contrast, maximum acceptable waste metal concentrations were defined for aerobically digested sludges and dewatered anaerobically digested sludges (Table 2). These sludges contain much less plant-available nitrogen than liquid anaerobically digested sludges but they contain substantial amounts of phosphorus, organic nitrogen, micronutrients and organic matter which are valuable to agriculture.

Canada and British Columbia. With the exception of Cd, Cr and Cu, the maximum acceptable waste metal loadings to soil in sludge represented for sale as fertilizers or soil supplements in Canada were chosen to approximate the Ontario values (Table 2). No limits were defined for Cr and Cu due to lack of evidence that buildup of these metals in soil represents a significant environmental risk. The limit for Cd was set at 4 rather than 1.6 kg/ha, based on considerable evidence (Ryan et al., 1982 and Davis and Coker, 1980) that the 1.6 kg/ha value is unnecessarily conservative. Maximum acceptable waste

Table 2. Waste Metal Guidelines for Agricultural Utilization of Municipal Sludge in Canada

	Maximum Acceptable Concentrations in Sludge (mg/kg dry wt)			Maximum Acceptable Loadings to Soil (kg/ha)								Maximum Acceptable Concentrations in Soil (mg/kg dry wt)	
	Canada[a]	Ontario[b]	British Columbia[c]	Quebec[d]	Canada	Ontario	British Columbia[d]	Alberta	Saskatchewan	Manitoba[f]	Nova Scotia	Quebec	Ontario
As	75	170	75	10 (20)	15	14	15		12	0.85	9	7.5	14
B				100 (200)				5-10		580			
Cd	20	34	20 (25)	10 (15)	4	1.6	4	0.8-1.5	1.2	1.1	1	2	1.6
Co	150	340	150	50 (100)	30	30	30		30	1.4	19	15	20
Cr		2800		500 (1000)		210		50-100	100	205	130		120
Cu		1700		600 (1000)		150		100-200	120	45	95		100
Hg	5	11	5 (10)	5 (10)	1	0.8	1	0.2-0.5	0.4	0.15	0.5	0.5	0.5
Mn				800 (1500)									
Mo	20	94	20	20 (25)	4	4	4		4	< 2.8	2.5	2	4
Ni	180	420	180 (200)	100 (180)	36	32	36	12-25	30	4.8	20	18	32
Pb	500	1100	500 (1000)	300 (500)	100	90	100	50-100	80	60	55	50	60
Se	14	34	14	14 (25)	2.8	2.4	2.8		2	0.15	1.5	1.4	1.6
Zn	1850	4200	1850 (2500)	1750 (2500)	370	330	370	150-300	300	225	200	185	220

[a] Values are for sludge and sludge-based products containing $\leq 5\%$ N dry wt basis and represented for sale as fertilizers or soil supplements.

[b] Values are for aerobically digested sludges and dewatered anaerobically digested sludges.

[c] Values are for agricultural high grade and, in parentheses, low grade sludges.

[d] Values are recommended maxima and, in parentheses, obligatory (i.e., not to be exceeded) concentrations.

[e] Choice of value depends upon site suitability for sludge application.

[f] Values derived assuming 56 tonnes/ha dry wt of Winnipeg sludge applied to land and the 1980 mean waste metal concentrations in sludge as follows: As, 15; B, 10 400; Cd, 19.5; Co, 25; Cr, 3700; Cu, 770; Hg, 2.6; Mo, <50; Ni, 85; Pb, 1050; Se, 2.6; Zn, 4050 mg/kg dry wt (Manitoba brief, 1981).

metal concentrations in sludge (Table 2) were calculated assuming the maximum acceptable waste metal loadings to soil and a 200 t/ha dry wt cumulative sludge loading. The Canadian limits were adopted for the draft British Columbia Guidelines.

Alberta. The Alberta limits for waste metal loadings to soil (Table 2) vary depending upon site suitability for sludge application. Sites are classified according to soil pH, texture, slope and depth to groundwater (Table 3). The metal loading limits increase with increasing site suitability and parallel but are more conservative than the Ontario limits. Agricultural land in Alberta is easily available for sludge spreading and it was decided that waste metal buildup in soil should be minimized.

Table 3. Classification of Sludge Application Sites in Alberta
(McCoy et al., 1982)

	Suitable for Sludge Application			Not Suitable for Sludge Application
	Class 1	Class 2	Class 3	Class 4
pH	≥6.5	≥6.5	≥6.5	<6.5
Texture^a	CL,SiCL,SiL, Si,SiC,L,SCL SC	C,HC	LS,SL	Sand and gravel
Slope %	0-2	2-5	5-9	>9
Depth to Potable Aquifer (m)	>5	3-5	2-3	<2

^{a}C-clay, L-loam, Si-silt, S-sand and H-heavy

Saskatchewan. Maximum acceptable waste metal loadings to soil (Table 2) defined in recent draft guidelines for sludge use in agriculture are generally smaller than the Ontario values.

Manitoba. In Manitoba, agricultural utilization of Winnipeg sludge is regulated by a Clean Environment Commission Order. Currently, land may receive one 56 tonne/ha application of sludge and the maximum acceptable waste metal loadings to soil (Table 2) were derived assuming 1980 mean metal concentrations for the sludge (Manitoba brief, 1981).

Quebec. Recommended and obligatory (not to be exceeded) waste metal concentrations in sludge for agricultural utilization, have been defined (Table 2). Official approval and soil and plant monitoring are required for utilization of sludges exceeding the recommended values. The maximum acceptable waste metal concentrations for Quebec soils were derived from the Canadian "maximum acceptable loadings to soil" (Table 2). It was assumed that the weight of one hectare of soil, 15 cm deep is two million kg and deep, is two million kg and that, for example, adding 15 kg/ha As would result in a 7.5 mg/kg As concentration in soil. In general, the maximum acceptable waste metal concentrations for Quebec soils are smaller than for Ontario soils.

Nova Scotia. Guidelines were developed to regulate agricultural utilization of sludge in Pictou County. In general, the limits for waste metal loadings to soil are similar to the Alberta maxima.

Canadian Research on Heavy Metal Uptake

Canadian research on heavy metal uptake from sludge treated soil has been conducted in the greenhouse, in lysimeters, and in the field. However, experience has shown that greenhouse data may be misleading because uptakes are larger than in lysimeters or the field. The following is an abstract of lysimeter and field experimental information.

Yields of orchard grass, bromegrass, wheat and corn grown in Ontario following land application of sludge were as large as, or larger than, following application of commercial fertilizer (Webber et al., 1981). There was no evidence of yield reduction from sludge even where applications greatly exceeded the recommended rate (Ontario, 1986). Cadmium loadings in sludge ranging from 4 to 7.4 kg/ha caused minor increases in the Cd concentrations in orchard grass, bromegrass, wheat straw and grain and corn grain. However, the Cd concentration in corn stover increased from approximately 0.4 to 1 mg/kg dry wt. Zinc loadings ranging from 341 to 1032 kg/ha caused small increases in the Zn concentrations in orchard grass, bromegrass, wheat straw, wheat grain and corn grain. The Zn concentrations in corn stover increased from approximately 40 to 160 mg/kg. The Cd and Zn concentrations, respectively, in corn stover were related to the amounts of nitrilotriacetic acid-extractable Cd and Zn in soil and to the cumulative amounts of Cd and Zn added to soil in sludge (Soon et al., 1980). In general, including soil pH as an independent variable in regression analysis improved the above-mentioned relationships. Copper loadings ranging from 240 to 320 kg/ha caused approximately 2 mg/kg dry wt increases in the Cu concentrations of all plant materials except corn grain which exhibited no increase. Nickel loadings of 910 and 624 kg/ha, respectively, increased the Ni concentrations in orchard grass from approximately 2 to 21 mg/kg and of bromegrass from approximately 1 to 6 mg/kg. A loading of 536 kg/ha increased the Ni concentration in corn stover and grain approximately 1 mg/kg. Lead and Cr loadings up to 450 and 736 kg/ha, respectively, did not significantly increase the concentrations of these metals in crops.

The Cd and Zn concentrations in plant materials (mainly corn leaves) taken from old sludge application sites in Ontario, generally exhibited small increases over those in control materials (Webber et al., 1983). Increases for Cd ranged up to 0.7 mg/kg with a median value of 0.02 mg/kg, and for Zn up to 100 mg/kg with a median value of 7 mg/kg. There was no consistent effect of sludge application on the Cu, Ni, Pb, Mo (except at Stratford), and Cr concentrations in plant materials. The Stratford sludge added much more Mo to soil than the other sludges and increased the Mo concentration in corn leaves from 0.7 to 32 mg/kg. The largest estimated metal loadings to these sites were Cd, 3.1; Zn, 595; Cu, 111, Ni, 118; Pb, 90; Mo, 19.4; and Cr, 628 kg/ha.

Wheat grain and straw grown on a calcareous soil treated with 400 to 800 tonnes/ha of Winnipeg sludge exhibited no increase in Pb concentration, minor increases in Cd and Cu concentrations and moderate increases in Zn concentrations (Zwarich and Mills, 1979). With one exception each, the Cd and Zn concentrations were <0.5 mg/kg dry wt and <60 mg/kg dry wt, respectively. Estimated metal loadings to the soil assuming the following sludge composition: Cd, 18; Zn, 2060, Cu, 495; and Pb, 370 mg/kg dry wt and a 600 tonnes dry wt/ha sludge application rate were: Cd, 11; Zn, 1240; Cu, 300; and Pb, 220 kg/ha.

In Alberta, sludge application generally increased yields of alfalfa grown on a solonetzic soil at Lethbridge (Graveland, 1983a) and on a calcareous chernozem soil at Edmonton (Lutwick and Graveland, 1983), and of barley grown on a calcareous chernozem soil at Calgary (Graveland, 1983b). However, the effects of sludge application on heavy metal concentrations in the plant materials were minor. Alfalfa exhibited occasional increases in Zn and small decreases in Mn. Barley straw exhibited increases in Zn, occasional increases in Pb and small decreases in Mn. Barley grain exhibited occasional increases in Cd and Zn. Frequently, the differences in concentrations between years greatly exceeded the differences due to sludge treatment. Copper, Ni and Cr concentrations in the plant materials were not affected by sludge treatment. The Calgary, Edmonton and

Lethbridge sludges exhibited low to moderate waste metal concentrations and although the maximum sludge loadings to experimental sites ranged from 30 to 78 tonnes dry wt/ha, the maximum metal loadings were only: Cd, 0.7; Zn, 53; Cu, 28; Ni, 4; Pb, 48; Cr, 75; and Mn 19 kg/ha.

Future Agricultural Utilization

Agricultural utilization of sludge is a well established practice in Canada and is usually much cheaper than disposal where there is relatively easy access to land. It accounts for twenty-nine percent of the estimated total sludge production and forty-four percent of the estimated sludge removal from wastewater treatment facilities and is expected to remain an important future sludge management option.

The risks associated with waste metal application to agricultural land are well recognized and guidelines to limit loadings to soil, have been developed for several Canadian jurisdictions. There is variability between the limits (Table 2), however, they generally correspond with the mid-range of values adopted by a large number of other countries (Table 4). Canadian research on waste metal uptake by plants from sludge treated soils indicates that even the maximum suggested values are not likely to cause significant crop production or animal and human health problems. Further research and practical experience may indicate that some relaxation of limits is warranted.

Agricultural utilization of sludge according to the Ontario guidelines has been practiced for several years and has proven satisfactory for both the agricultural and wastewater treatment communities. The guidelines have prompted considerable reductions of waste metal concentrations in several sludges. These reductions were necessary to maintain sludge acceptability for agricultural utilization and in most cases were accomplished by improved management of industrial processing or pretreatment of industrial effluents. Only a few sludges have been declared unacceptable for agricultural utilization due to high metal concentrations.

Although agricultural utilization of sludge in Canada is based on sound scientific information there are pockets of public resistance usually due to incorrect information or unsatisfactory past experience. Efforts to increase public acceptance have included the development of two information packages (Black et al., 1984 and Webber, 1984) and audio-visual presentations.

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Table 4. Waste Metal Guidelines for Agricultural Utilization of Sludge (Webber et al., 1984)

	Maximum Acceptable Loadings to Soil (kg/ha)											
											All Countries	
	Canada	Denmark	Finland	France	Germany	Netherlands	Norway	Sweden*	United Kingdom	United States	Range	Median
As	15					2			10		2-15	10
Cd	4	0.2	0.1	5.4	8.4	2	0.2	0.075	5	5, 10, 20	0.1-20	5
Co	30						0.4	0.25			0.4-30	
Cr				360	210	100	4	5	1000		4-1000	210
Cu				210	210	120	30	15	280	125, 250, 500	30-500	210
F									600			
Hg	1			2.7	5.7	2	0.14	0.04	2		0.14-5.7	2
Mn							10					
Mo	4								4			
Ni	36			60	60	20	2	2.5	70	50, 100, 200	2-200	60
Pb	100			210	210	100	6	1.5	1000	500, 1000, 2000	6-2000	210
Se	2.8								5		2.8-5	
Zn	370			750	750	400	60	50	560	250, 500, 1000	60-1000	500

* Five-year loadings can be repeated. Values are not included in "All Countries Range and Median".

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